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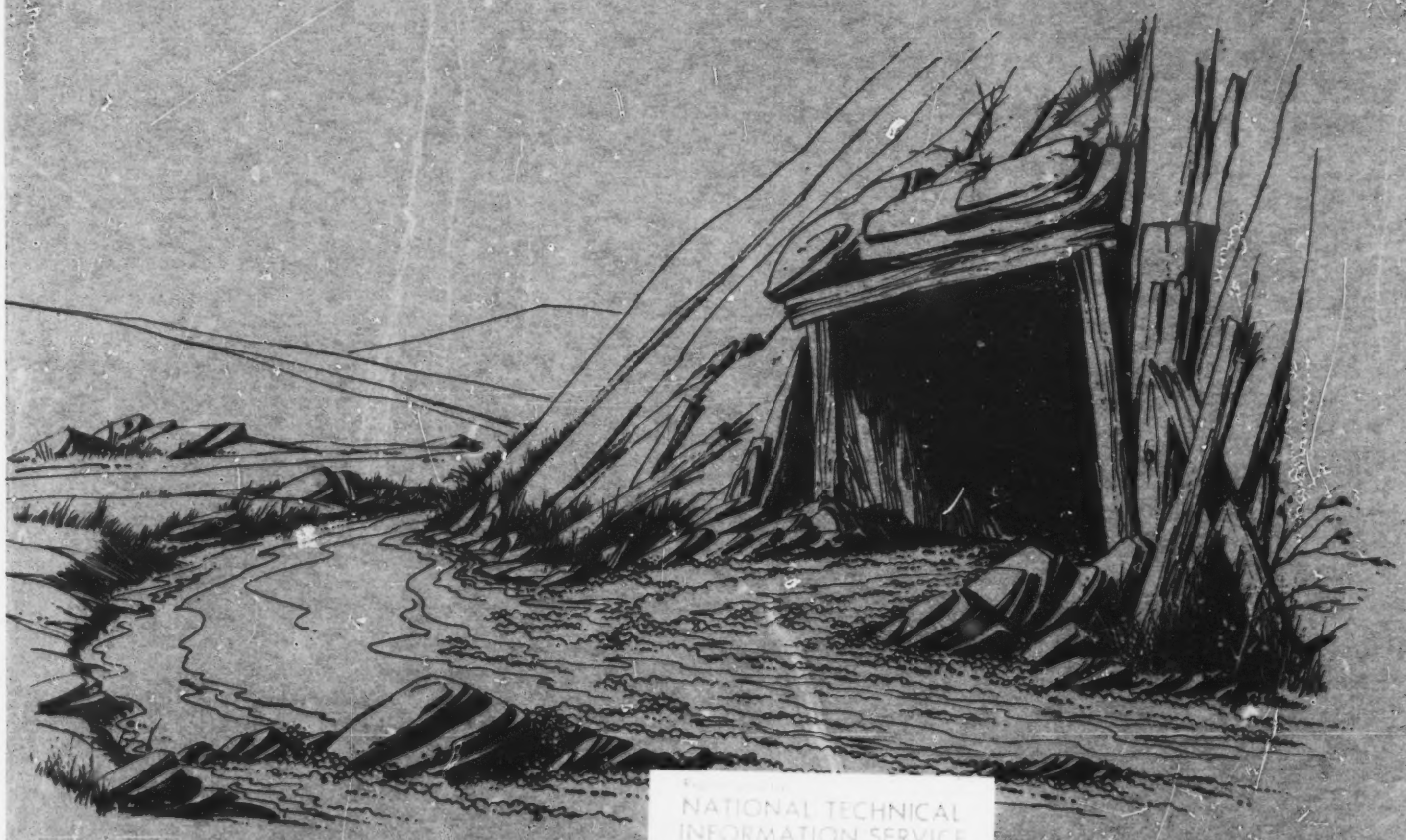
REVERSE OSMOSIS DEMINERALIZATION OF ACID MINE DRAINAGE



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REVERSE OSMOSIS DEMINERALIZATION OF ACID MINE DRAINAGE

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A two-phase study, involving both laboratory and field investigations, has demonstrated the feasibility of using reverse osmosis to provide potable water from acid mine drainage. The laboratory investigations involved the determination of methods for controlling iron fouling and the selection of a process flow sheet. During the field test, the process developed in Phase I was used to treat acid mine drainage from an underground abandoned anthracite coal mine. Treatment prior to reverse osmosis consisted of filtration (10) followed by ultraviolet light disinfection. Brine from the RO unit was treated by neutralization, oxidation, and settling. Results obtained indicated that membrane fouling due to iron was satisfactorily controlled, but calcium sulfate fouling limited the recovery of product water to about 75%. Product water was of potable quality in all respects except for iron, manganese, and pH. Rejection of individual ions across the RO membranes ranged from 99.2 to 99.7% based on average brine concentrations. Product waters of 25 mg/l total dissolved solids were produced from a feed TDS level of 1319 mg/l. Calcium sulfate precipitate on the RO membrane was successfully removed using a solution of ammoniated citric acid at pH 8. (Lowry-Texas)

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REVERSE OSMOSIS DEMINERALIZATION

OF ACID MINE DRAINAGE

by

The Ecology Division
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for the

Commonwealth of Pennsylvania
Department of Environmental Resources
Harrisburg, Pennsylvania 17102

and the

Office of Research and Monitoring
Environmental Protection Agency

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ABSTRACT

The objective of this study was to determine the operational methods and procedures necessary to successfully demineralize acid mine drainage utilizing reverse osmosis (RO). The study was conducted in two phases. Phase I consisted of laboratory bench scale investigations to determine methods for controlling iron fouling and to select a process flow sheet. Phase II was the field operation based on the flow sheet selected in Phase I.

The field test site was located in Mocanaqua, Pennsylvania. The source of acid mine drainage was the discharge from an abandoned underground anthracite coal mine. Treatment prior to RO consisted of filtration (10μ) followed by ultraviolet light disinfection. The brine from the RO unit was treated by neutralization, oxidation and settling. The field test phase spanned a four month period. Frequent samples were analyzed to characterize the operation of the system.

The results obtained indicated that it was feasible to demineralize acid mine drainage by reverse osmosis. Membrane fouling due to iron was satisfactorily controlled. The recovery of product water was limited to about 75% due to calcium sulfate fouling. Product water was of potable quality in all respects except for iron, manganese, and pH. Neutralization, oxidation and filtration would be required to meet potable standards.

This study was performed under Contract No. CR-86-A with the Commonwealth of Pennsylvania, Department of Environmental Resources. Partial sponsorship was provided by EPA (Program No. 14010 FQR).

CONTENTS

Section		Page
I	CONCLUSIONS	1
II	RECOMMENDATIONS	3
III	INTRODUCTION	5
IV	LITERATURE SEARCH	7
V	LABORATORY INVESTIGATIONS	9
	Iron Fouling Investigations	15
	Investigation of Alternate Flow Schemes	20
VI	DESCRIPTION OF FIELD DEMONSTRATION SYSTEM	35
VII	DISCUSSION OF FIELD TEST RESULTS	43
	Raw AMD Characteristics	43
	Operation of the Pretreatment System	47
	Operation of the Tubular RO System	50
	Operation of the Hollow Fiber RO System	57
	Operation of the Brine Treatment System	66
VIII	GENERAL DISCUSSION	73
	Discussion of Flushing Techniques	73
	Economic Consideration for RO-AMD Operation	74
IX	ACKNOWLEDGEMENTS	79
X	REFERENCES	81
XI	PUBLICATIONS	83
XII	GLOSSARY OF TERMS	85
XIV	APPENDICIES	87
	I. Operating Data Hollow Fiber RO System	87
	II. Tubular RO System Data	104

FIGURES

		Page
1	TUBULAR RO MODULE CONFIGURATION	10
2	SPIRAL WOUND CONFIGURATION	12
3	HOLLOW FIBER MODULE	L3
4	SCHEMATIC FLOW DIAGRAM FOR LABORATORY RO UNIT	14
5	IRON FOULING INVESTIGATIONS SIMULATED AMD	16
6	IRON FOULING INVESTIGATIONS (Type 300 Tubular Modules)	19
7	ALTERNATE AMD TREATMENT SCHEME - METHOD A, LIMITED PRETREATMENT	21
8	ALTERNATE AMD TREATMENT SCHEME - METHOD B, PRENEUTRALIZATION	27
9	OPERATIONAL DATA FOR PRENEUTRALIZATION (METHOD B)	28
10	ALTERNATE AMD TREATMENT SCHEME - METHOD C, PRENEUTRALIZATION & SETTLING	32
11	SCHEMATIC DIAGRAM OF THE FIELD TEST APPARATUS	36
12	FIELD TEST APPARATUS	37
13	INITIAL MODULE ARRANGEMENT TUBULAR RO SYSTEM	38
14	MODULE ARRANGEMENT HOLLOW FIBER RO SYSTEM	40
15	HOLLOW FIBER MODULES	41
16	NEUTRALIZATION SYSTEM	42
17	RAW WATER QUALITY VARIATION	48
18	IRON OXIDATION RATES	49
19	MODULE ARRANGEMENTS UTILIZED FOR TUBULAR 310 MODULES	51
20	TUBULAR RO SYSTEM OPERATION WITH TYPE 31- MODULES	53
21	TUBULAR RO SYSTEM OPERATION WITH TYPE 610 MODULES	
22	HOLLOW FIBER RO OPERATION PERMEATOR #691	58
23	HOLLOW FIBER RO SYSTEM OPERATION	59
24	COMPARISON OF $CaSO_4$ MOLAR SOLUBILITY PRODUCT AND PRODUCT WATER FLOW	63
25	IRON OXIDATION STUDY	67
26	SETTLING RATE TESTS	68
27	FLOW SHEET USED FOR COST ESTIMATES	76

TABLES

No.		Page
1	TYPICAL COMPOSITION OF SIMULATED ACID MINE WATER	15
2	SUMMARY OF THE OPERATING CONDITIONS FOR RO EXPERIMENTS IN THE LABORATORY	17
3	DATA SUMMARY ON SIMULATED AMD	22
4	CHANGE IN BRINE CHARACTERISTICS WITH TIME	24
5	CHEMICAL ANALYSIS OF HOLLOW FIBER RO BRINE	25
6	ANALYSIS OF SOFTENED RO BRINE	25
7	TYPICAL WATER QUALITY DATA FOR THE IRON SLURRY EXPERIMENT	29
8	MANGANESE REMOVAL WITHOUT pH ADJUSTMENT	30
9	OPERATIONAL DATA FOR SCHEME C	33
10	RAW ACID MINE DRAINAGE CHARACTERISTICS	44
11	COMPARISON OF FIELD AND LABORATORY ANALYSIS OF RAW ACID MINE DRAINAGE	45
12	SALT REJECTION CHARACTERISTICS TUBULAR RO SYSTEM	56
13	SALT REJECTION CHARACTERISTICS HOLLOW FIBER RO SYSTEM	65
14	BENCH SCALE NEUTRALIZATION TESTS	70
15	SUMMARY OF NEUTRALIZATION SYSTEM OPERATION	71
16	SUMMARY OF ATOMIC ABSORPTION ANALYSIS OF NEUTRALIZATION SYSTEM	72
17	LABORATORY RO MODULE CLEANING RESULTS	75
I-1	HOLLOW FIBER OPERATING DATA FOR SINGLE PERMEATOR	88
I-2	HOLLOW FIBER OPERATING DATA 2-1 ARRAY	90
I-3	OPERATIONAL DATA HOLLOW FIBER RO SYSTEM (2-1 ARRAY FIRST STAGE MODULES)	93
I-4	HOLLOW FIBER SYSTEM OPERATION (2-1 ARRAY, 2ND STAGE MODULE)	96
I-5	ANALYSIS DATA HOLLOW FIBER SYSTEM (AA DATA)	99
I-6	ANALYSIS DATA HOLLOW FIBER SYSTEM (AA DATA)	100
I-7	FIELD ANALYSIS DATA HOLLOW FIBER SYSTEM	101
I-8	FIELD ANALYSIS DATA HOLLOW FIBER SYSTEM	102
I-9	FIELD ANALYSIS DATA HOLLOW FIBER SYSTEM	103
II-1	TUBULAR RO SYSTEM OPERATIONAL DATA 310 MODULES	105
II-2	TUBULAR RO OPERATIONAL DATA 610 MODULES	108
II-3	FIELD ANALYSIS DATA TUBULAR RO SYSTEM	109
II-4	LABORATORY ANALYSIS DATA TUBULAR RO UNIT	110

SECTION I

CONCLUSIONS

The following conclusions were made based on the data obtained during this study:

1. The feasibility of using Reverse Osmosis (RO) to provide potable water from acid mine drainage was demonstrated.

2. The flux declines observed were tolerable (slope of log-log plot of flux and operating time less than 0.031) and flux rates can be sustained with a minimum of membrane flushing.

3. Oxidation of iron (II) by bacteria can be controlled by ultraviolet light disinfection or lowering the pH of the feed water.

4. The acid mine drainage should not be neutralized prior to RO processing.

5. Feed water pH is critical with regard to iron fouling of RO membranes. Iron fouling can be controlled completely at a pH of ~2.8 or below.

6. Allowable product water recovery is strongly influenced by the $CaSO_4$ concentrations in the brine.

7. Calcium sulfate fouling of the RO membranes was found to occur above a $CaSO_4$ molar solubility product of 25 to 35 x 10^{-5}, as measured in the brine stream.

8. Rejection of individual ions across the RO membranes was in the range of 99.2 to 99.7 percent based on average brine concentrations.

9. Product waters of 25 mg/l total dissolved solids (TDS) were produced from a feed TDS level of 1319 mg/l.

10. The product waters produced did not meet the USPHS recommendations for iron, manganese, and pH, hence, would require limited additional treatment.

11. A high flux decline rate was observed for the tubular system when low salt rejection (98.5%) membranes were utilized, while the higher salt rejection (99.6%) membranes had significantly improved flux stability.

12. Iron oxide precipitation on the RO membranes was successfully removed using a solution of sodium hydrosulfite ($Na_2S_2O_4$).

13. Calcium sulfate precipitation on the RO membranes was successfully removed using a solution of ammoniated citric acid at pH 8.

14. No damage to the RO membrane desalting properties was observed due to 2670 hours of sustained operation on acid mine drainage or the various flushing solutions utilized.

15. Neutralization of the brine to a pH of 7.9, followed by oxidation and settling did not produce an effluent which could be reprocessed by the RO system.

SECTION II

RECOMMENDATIONS

1. Additional studies on RO brine treatment be made to determine the necessary treatment which will allow the brine to be recycled and thus eliminate this waste disposal problem.

2. Studies should be initiated to investigate recovery of iron and aluminum for use as coagulants at sewage treatment plants. Successful recovery could lead to reduced operating costs.

3. In acid mine drainage where the majority of the iron is in the form of iron (II), a small amount of iron (III) generally exists. The role of this iron (III) with regard to membrane fouling should be further evaluated.

SECTION III

INTRODUCTION

The pollutional effects of acid mine drainage (AMD) as well as various methods of abating this pollution have been well documented as a result of federally sponsored projects (1)(2). Among the various pollution abatement techniques proposed, the use of reverse osmosis (RO) to purify AMD appears promising and has been under study since 1966. Most of the previous studies have been of short duration, that is less than 1000 hours continuous operation, and many have been less than 200 hours operation. These short term tests were not sufficient to identify the various operating problems which could occur in treatment AMD utilizing reverse osmosis. Furthermore, specific problems developed from certain studies regarding iron fouling (3) and calcium sulfate fouling (4) which required additional study to optimize the flow sheet and operating procedures for an AMD/RO treatment system.

The objectives of this study were:

1. Determine the causes of iron fouling previously encountered (3) and formulate methods of controlling and/or eliminating this type of fouling.

2. Investigate various alternate flow schemes for treating AMD utilizing RO.

3. Operate an AMD/RO treatment system for a sufficiently long operational period to establish reliable operating characteristics.

To accomplish the stated objectives, the project was divided basically into two phases, 1) a laboratory investigation phase to evaluate the iron fouling problem and investigation of various possible alternate flow schemes, 2) a field evaluation phase to operate the selected flow scheme for a period of 2400 continuous operating hours, evaluating such parameters as water recovery rates, rates of membrane fouling, permeate water qualities, specific operating procedures required to minimize membrane fouling, and membrane cleaning techniques.

The source of acid mine drainage (AMD) was the Mocanaqua discharge in Mocanaqua, Pennsylvania. This is the same discharge utilized in the previous study (3).

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SECTION IV

LITERATURE SEARCH

The feasibility of utilizing reverse osmosis (RO) to recover high purity product water from acid mine drainage (AMD) waters and to abate pollution has been under investigation since 1966 (5). Riedinger and Schultz (5) found that high quality water could be produced from acid mine drainage via reverse osmosis. The membrane system which was utilized was a spiral wound system marketed by Gulf Environmental Systems. Feed water pH was 3 or less and contained approximately 100 mg/l of iron. Water recoveries in excess of 90% were reported, but some iron fouling of the membrane was found to occur, decreasing the product water output. Other investigations had also indicated problems with iron fouling of RO membranes and it appeared that iron fouling and subsequent membrane cleaning was the most critical problem encountered in applying this process to the treatment of acid mine waters.

Hill (9), however, reporting on the work with the acid mine waters in Norton, West Virignia, indicated that no problems with iron fouling were experienced. Salt rejections as high as 99% were reported. The majority of the iron at this site was in the trivalent state. Kremen et al. (4) reporting from work at the same site, concluded that reverse osmosis could process acid mine drainage feed streams to high degrees of recovery, could produce excellent permeate water, and posed no special or difficult problems for reverse osmosis processing. They further stated that membrane lifetimes had been demonstrated which permitted confident cost projections for immediate technology and for reasonably certain near future state of the art. Sustained reverse osmosis operation up to 75% recovery levels were reported. Increased recovery levels up to 92% for short periods did not show the anticipated difficulties with calcium sulfate precipitation. Although fouling of the membranes at the discharge end of the plant did occur, these calcium sulfate scales could be removed by operating the unit at 50% recovery for short periods of time, thus flushing them from the membranes.

At the same time Mason (3), reporting on the work done at Shickshinny, Pennsylvania (Mocanaqua discharge), concluded that although a high quality product water could be produced via reverse osmosis, a number of operational problems needed to be investigated before RO could be applied to treat acid mine drainage on a large scale. This work was conducted on acid feed waters containing the majority of iron in the ferrous state. A tubular RO configuration manufactured by Calgon-Havens Industries was utilized for this study. The main problem emerging from this study was the maintenance of high water permeation rates due to membrane fouling by iron. It was also indicated that to utilize the permeates for potable use further treatment would be required when the iron content in the feed water exceeded 100 mg/l.

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The sustained membrane performance tests had so far been limited to water recovery rates below 80% because of the fear of fouling the membranes with calcium sulfate. A study (8) carried out at three different mine drainage sites by Gulf Environmental Systems under the sponsorship of EPA, concluded that the limiting factor in achieving the maximum water recoveries was the calcium sulfate concentration. No iron fouling was reported during operation of any of the three sites investigated. To further increase the water recovery rates, a combination of neutralization and reverse osmosis called the 'Neutralosis' process was proposed by the EPA staff (9). The process utilized the operation of the reverse osmosis unit at maximum recovery (~90%). The brine was then neutralized and settled and the overflow from the settling tank returned to the RO unit for reprocessing. It was concluded that the Neutralosis process produced 98% water recovery when operated on a predominantly ferric iron acid mine drainge. However, these results were based on relatively short term testing (less than 100 hours).

In view of the conflicting observations discussed above regarding iron fouling of the cellulose acetate membranes and current developments in the reverse osmosis membranes and hardware technology, it was apparent that many technical areas required further investigation in order to successfully apply reverse osmosis to the treatment of acid mind drainge.

SECTION V

LABORATORY INVESTIGATIONS

The objectives of the laboratory investigations were twofold:

1. To investigate the mechanisms involved in the fouling of RO membranes by iron.

2. Evaluation of alternate flow schemes to determine the most desirable method of utilizing RO to treat acid mine drainge.

These investigations were conducted with pilot scale RO equipment using both synthesized and actual acid mine waters. Three types of commercially available RO systems i.e. tubular, spiral wound and hollow fine fiber were utilized during these investigations.

The tubular RO system was manufactured by Calgon-Havens. Both the older module (type 300 - as utilized in a previous study (3)) and the modified modules (type 310 and 510) were utilized. There are two main differences between the older and the modified modules. First, the method of interconnection of the individual tubes within each module was different. The tubes in the older module were connected by separate turn arounds at both ends while in the new module they were connected internally by means of an integral head (sealed by o-rings between the tubes and the head). Second, the fiberglass tubes in the new module were significantly stronger than the older tubes and were expected to have considerably better life. The new tubes had been strengthened by utilizing new manufacturing procedures. Each module consisted of 18 porous fiberglass tubes with an effective membrane area of 16.9 sq ft. The new modules were also equipped with turbulance promotors within each tube.

The turbulance promotor, called volume displacement rod (VDR), was a helically wound rod and was placed inside the individual tubes to minimize the concentration polarization effects by increasing the effective brine velocity through the tubes. A diagram of the new tubular module and the turbulence promotor is shown in Figure 1. Two types of tubular cellulose acetate membranes were utilized during the laboratory investigation phase. The membranes with higher flux rates and lower salt rejection were designated as type 300 or 310, while the membranes with comparatively lower flux rates and higher salt rejection were designated as type 500 or 510.

The two other types of RO equipment utilized in these investigations were the spiral round system (obtained from Gulf Environmental Systems) and the hollow fiber system (obtained from E.I. DuPont). Both these configurations have the advantage of a high membrane area to volume ratio compared to the tubular system. The spiral wound module was rolled about a center tube much like a scroll and uses a mesh spacer for the feed flow distribution. The membrane used was a newly developed

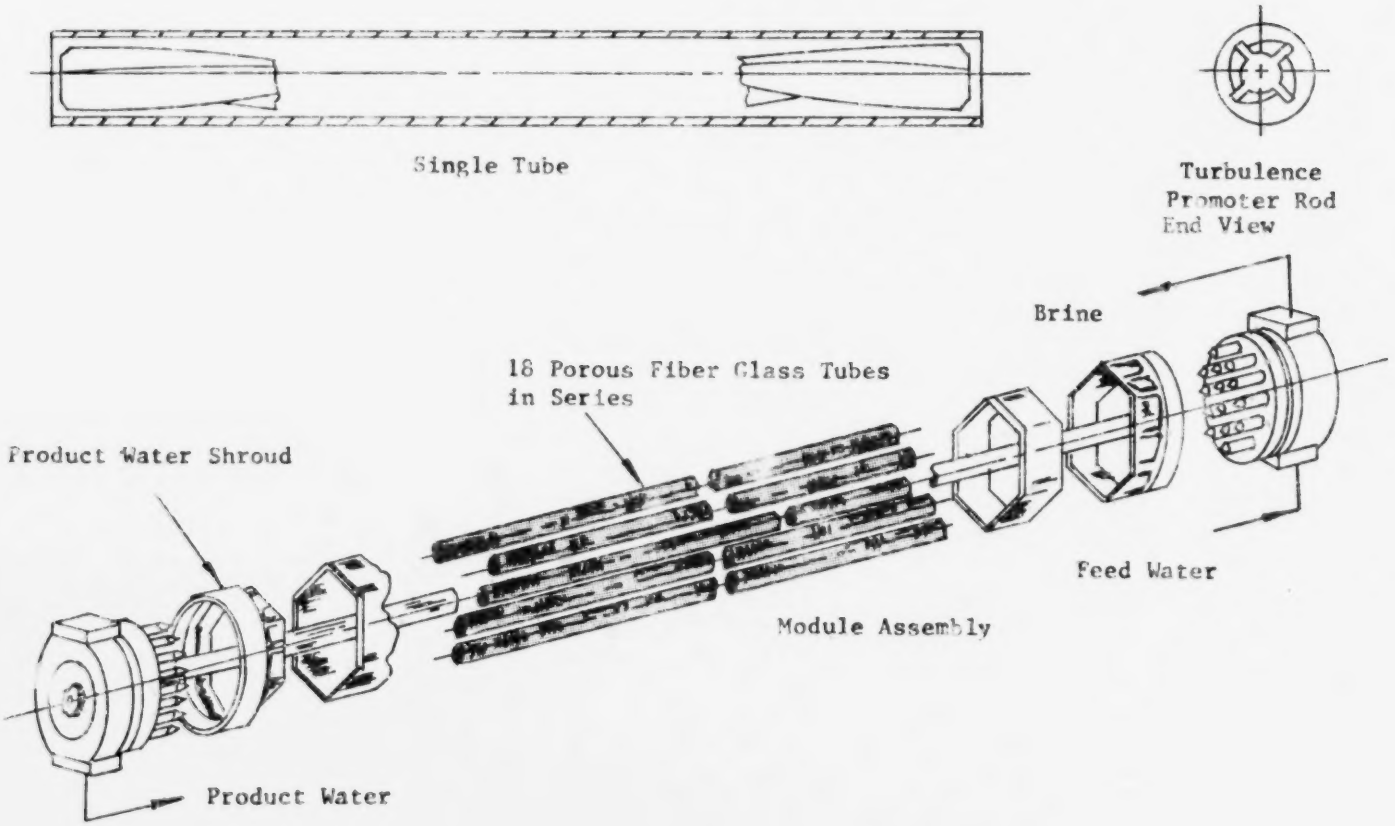

FIGURE 1

TUBULAR RO MODULE CONFIGURATION

high flux-high rejection, cellulose acetate membrane (Module Type 4001). The module in this system consists of one or more leaves wrapped around a product water take-off tube. These leaves consist of a membrane, porous incompressible product water side backing material, and a brine side flow spacer. The membrane is bonded along the two sides, at the end, and around the product water tube, forming a sealed envelope that encloses the backing material except at the product water tube open end. The brine side flow spacer is placed on the membrane, and several layers are then wrapped around the product water tube to form a cylindrical module. Modules are then placed in a pressure vessel which consists of a standard 4" schedule 40 steel pipe which has been coated for corrosion resistance. The pressure vessel utilized in this study was approximately 10 feet long and held 3 modules. Each module was three feet long and contained 50 sq ft of membrane surface. The product water tubes for each module are interconnected utilizing sleeves with "O" ring seals. Figure 2 presents a sketch of both the spiral wrapping configuration as well as the module arrangement within the pressure vessel.

The hollow fiber modules termed B-9 permeators utilized a newly developed polyamide membrane. This membrane is characterized with several advantageous feasures such as: significantly improved product water rates, lower operating pressures and higher salt rejection capabilities compared to the earlier membrane version 'B-5'. The B-9 module is 5.5 inches in diameter and 4 feet long. In each module, the individual hollow fibers (42 micron (u) inside diameter by 84u outside diameter) are bound into a cylindrical bundle containing a nominal fiber surface area of 1900 sq ft. The open ends are potted in epoxy to separate the purified water from the brine stream. This entire fiber assembly is installed in a tubular pressure vessel. The pressure vessel is normally made of aluminum. However, pressure vessels made of stainless steel or fiberglass have also been introduced by the manufacturer for special applications. Feed water under pressure circulates around the fibers. Pure water passes through the walls of the fibers and flows up the bore. The contaminants remain on the outside of the hollow fibers. The concentrate and permeate exit through separate outlets as shown in Figure 3.

A flow diagram of the pilot RO system utilized for the laboratory investigations is shown in Figure 4. The system consists of feeding the wastewater through the modules under high pressure by a Moyno pump. The pump speed is controlled by a variable drive and this controls the pump flow rate. A half inch diameter stainless steel coil using a recirculation of cold tap water was incorporated in the feed tank to control the feed water temperature in the range of 55 to 65°F.

In a typical experiment the pretreated wastewater was pumped to the membrane bank from the feed water tank. Both the concentrate and the permeate were recirculated to the feed tank. Measurements were recorded for TDS, temperature, pressure, pH and flow rates for the feed, concentrate and product streams. In order to simulate higher feed water recovery, the concentrate was continuously recirculated to the feed tank while the permeate was wasted until the desired recovery level was achieved.

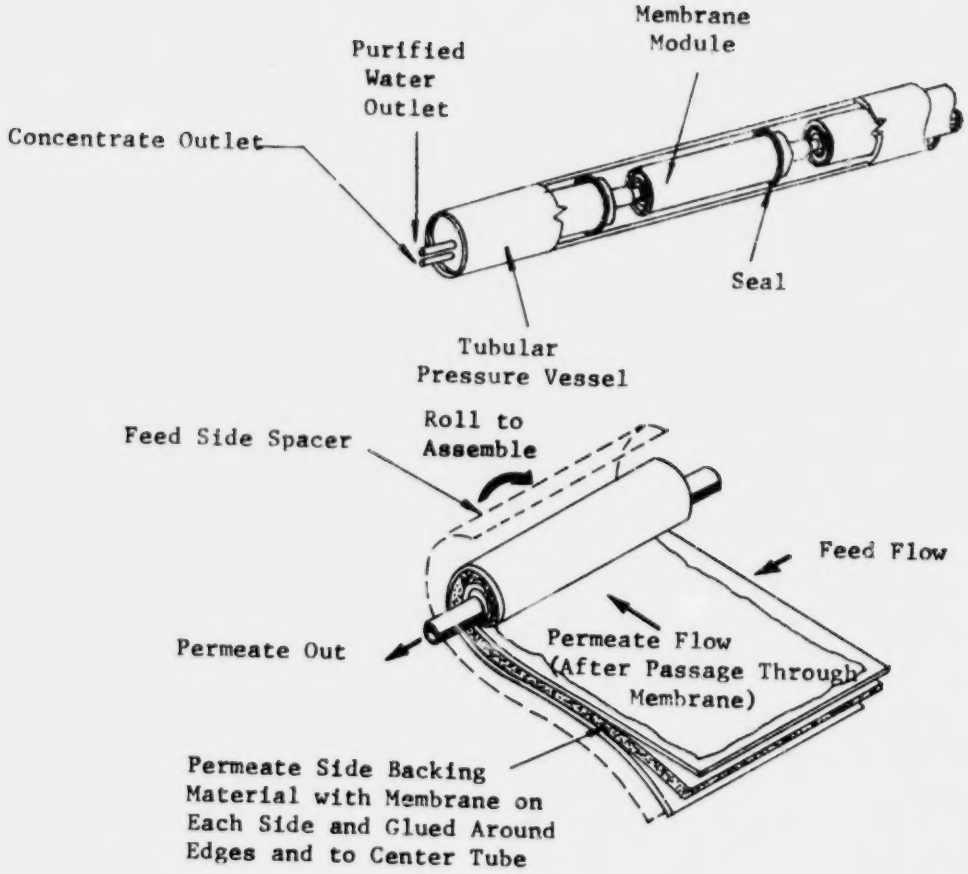

FIGURE 2

SPIRAL WOUND CONFIGURATION

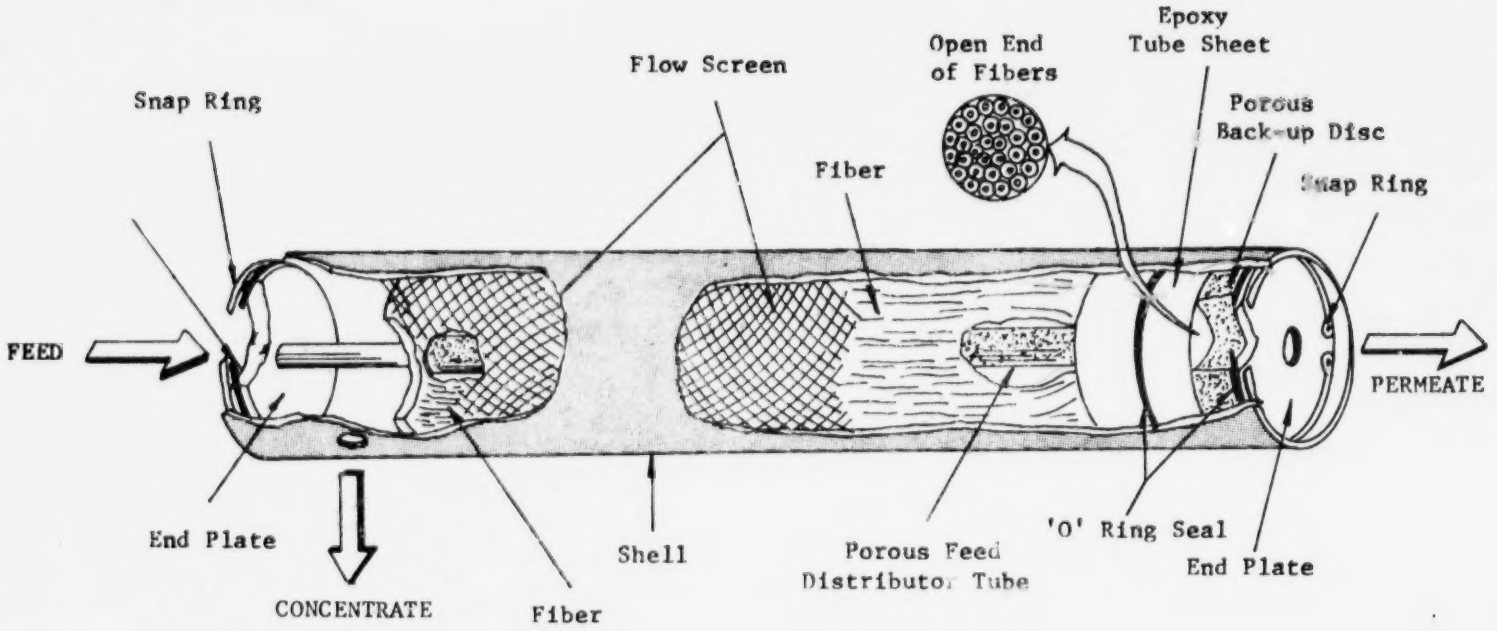

FIGURE 3

HOLLOW FIBER MODULE

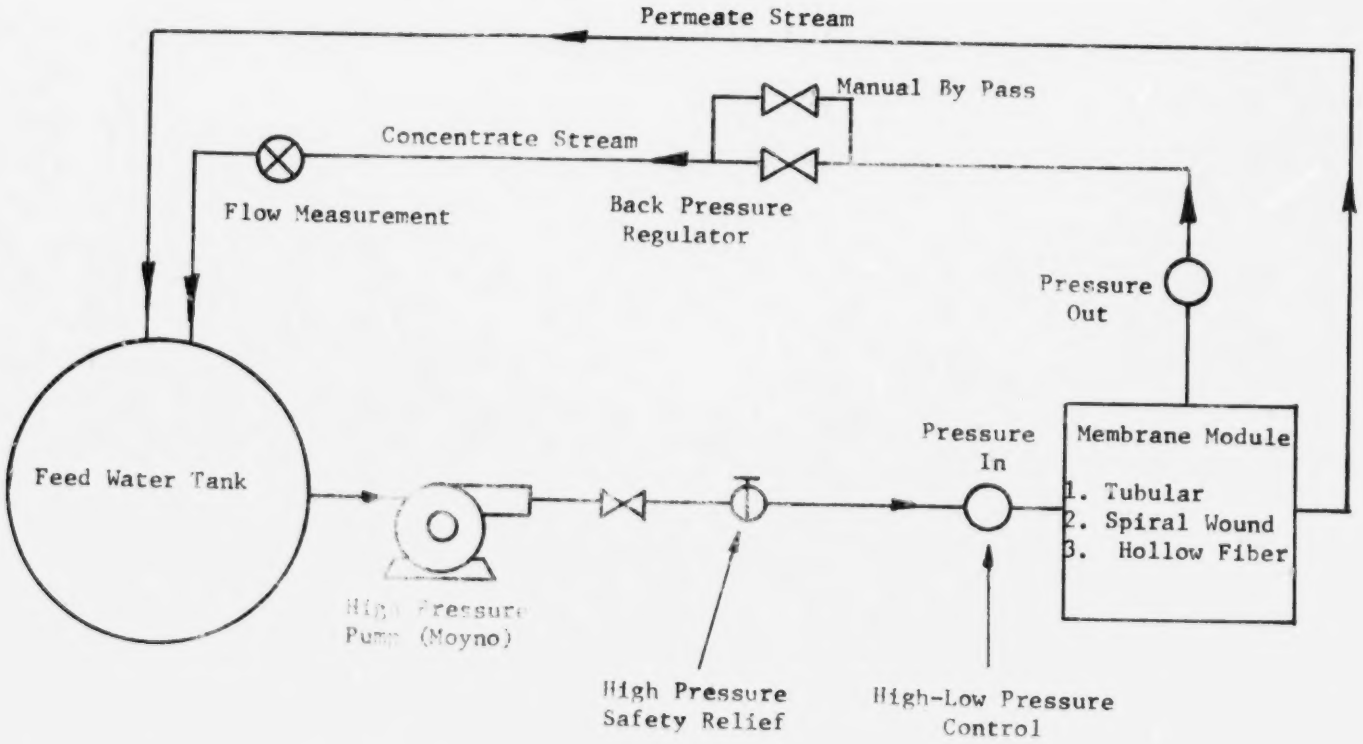

FIGURE 4

SCHEMATIC FLOW DIAGRAM FOR LABORATORY RO UNIT

The composition of the simulated acid mine water utilized during this study was similar to the one obtained for actual acid mine drainage at Shickshinny, Pennsylvania in an earlier study (3). Table 1 presents a typical composition of the simulated acid mine water. It should also be noted that these waste waters were simulated using Milwaukee tap water for all laboratory studies. A summary of the operating conditions for the three types of RO equipment utilized is shown in Table 2.

TABLE 1

TYPICAL COMPOSITION OF SIMULATED ACID MINE WATER*

Ion	mg/l
Calcium (Ca)	140
Magnesium (Mg)	100
Manganese	16
Iron (Fe^{++})	120
Sulfate (SO_4)	800
pH, units	3.6

*Based on the field analysis performed at Shickshinny, Pennsylvania during Fall, 1969.

Iron Fouling Investigations

The possible factors which could influence iron fouling of the RO membranes as observed in the field (3) were postulated to be as follows:

1. Chemical oxidation of iron (II) to iron (III) due to oxygen present in the feed water and subsequent precipiration of iron (III) compounds on the RO membranes.

2. Biological oxidation of iron (II) to iron (III) by iron oxidizing bacteria in the presence of oxygen and precipitatation of iron (II) compounds on the RO membranes.

3. The influence of concentration polarization on the rates of oxidation both chemical and biological

A total of three 250-300 hour laboratory tests were conducted with the tubular RO system to study the iron oxidation problem. Both simulated and actual AMD were utilized. The composition of the simulated waters duplicated as closely as possible the actual AMD characteristics (see description of test apparatus).

Experiment #1 was made with the simulated acid mine waters to evaluate the flux decline characteristics of the tubular membranes, type 310 and 510. Figure 5 presents the variation of flux rates for this experiment. The flux rate for membrane type 310 decreased 11% in the first 65 hours. and then remained steady until the end of the experiment (325 hours). The corresponding drop in flux rate for membrane type 510 was 15%. The

TABLE 2

SUMMARY OF THE OPERATING CONDITIONS FOR RO EXPERIMENTS IN THE LABORATORY

RO Configuration	Tubular	Spiral Wound	Hollow Fiber
Manufacturer	Calgon-Havens	Gulf Env.Sys.	DuPont
Membrane Type	300, 310 and 510	4001	B-9
Individual Module Size	3.5" dia. x 8'	4½" dia. x 3'	5.5" dia. x 4'
Membrane area per Module	16.0 sq ft	50 sq ft	1900 sq ft
Modules Utilized	3 to 4	3	1
Feed Pressure	600 psi	600 psi	400 psi
Feed Temperature	55 - 65°F	55 - 65°F	55 - 65°F
Water Output per Module	0.1 - 0.15 gpm	0.6 - 0.8 gpm	1.2 - 1.4 gpm
Minimum Brine Flow	0.75 gpm	3.0 gpm	0.5 gpm
Range of Feed Water pH	3.6 - 7.0	3.6 - 7.0	6.7
Dissolved Oxygen	7.8 - 8.5 mg/l	7.8 - 8.5 mg/l	7.8 - 8.5 mg/l
Feed Water Recoveries	25 - 60%	35 - 45%	40 - 85%

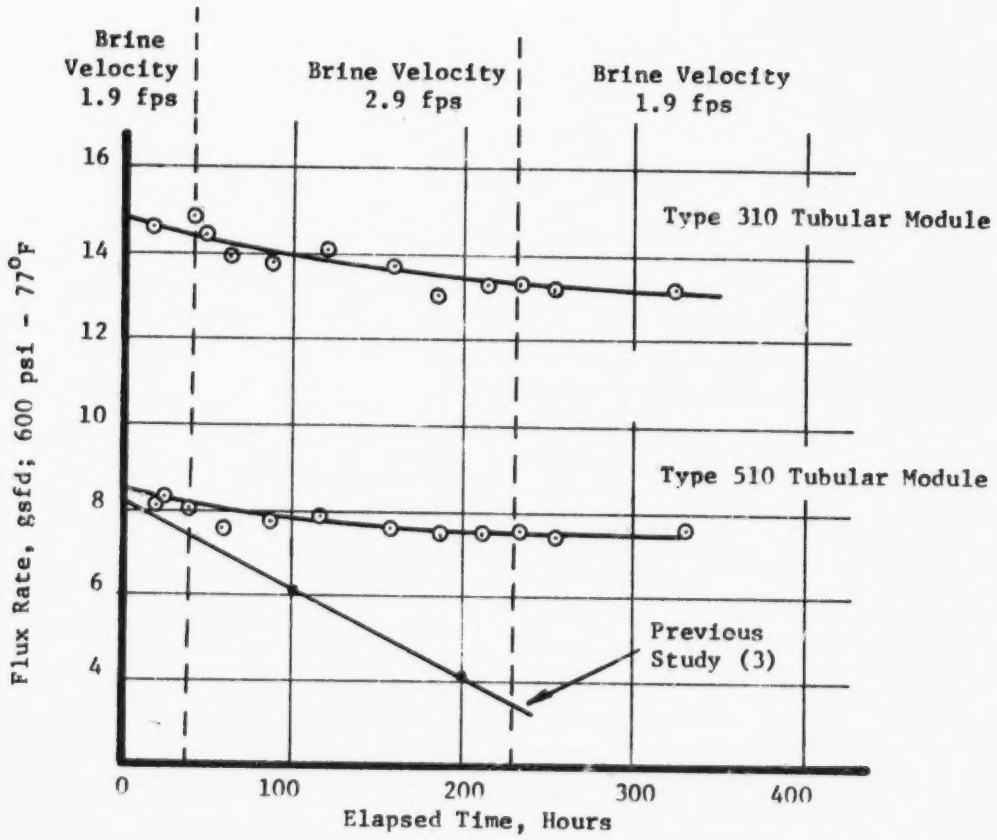

FIGURE 5

IRON FOULING INVESTIGATIONS SIMULATED AMD

reason for the initial drop in flux for membrane types 310 and 510 can apparently be attributed to compaction of the membrane at 600 psi. These declines could not have been due to iron fouling of the membranes as the flux rates stabilized after the initial decline. This observation was also confirmed by the consistent level of total iron and ferrous iron concentrations in the feed water at various intervals during the course of the experiment.

In an effort to study the effect of velocity on flux rates, the feed flow was increased from 0.75 gpm (1.9 fps) to 1.15 gpm (2.9 fps) after 45 hours of operation at the former velocity. At 118 hours of operation, the safety rupture disc failed and caused a brief shutdown of the system. As expected, partial restoration of the flux rates was noticed for both types of membranes (Figure 5) due to the depressurization effect. However, the flux rates indicated that the increase in velocity (meaning increased turbulence) had little influence on the decline of flux rates. To verify this observation, the feed velocity was reduced to the initial level of 1.9 fpm at 230 hours. No significant change in the flux rates was noticed in an additional 100 hours of operation. Therefore, it was concluded that there was no significant effect of velocity on the flux rates within the range investigated.

Comparing the flux decline rates for membrane type 310 from experiment #1 above to the rates attained in a previous field investigation (3), it was seen that the flux decline rates observed in the field were significantly greater than in the laboratory (see Figure 5).

Although the composition of the simulated AMD was similar to the Mocanaqua discharge and was also saturated with dissolved oxygen, other factors that could affect the flux characteristics were:

1. New tubular modules with strengthened tubes containing turbulence promoter rods.

2. Absence of an ambient atmosphere containing iron bacteria.

To investigate the effect of the above factors, two additional tests were performed with old Havens modules type 300. Experiment #2 was a duplication of experiment #1 with the exception of utilizing the old Havens without turbulence promoters. Experiment #3 was conducted with a 4:1 combination of the simulated acid mine water and actual Mocanaqua discharge to provide some iron bacteria in the laboratory test solution.

Figure 6 shows a comparison of the flux decline characteristics for the experiments 2 and 3 under the present study as well as for the field investigation during the fall of 1969 (3). Comparing the flux decline curves for the membrane type 300 with or without the iron bacteria (Experiments #2 and #3 respectively), the flux rate characteristics were found to be quite similar. It could therefore be concluded that the presence of the iron bacteria did not have any significant effect on the flux rates. However, such a conclusion might not be justified in light of the following factors:

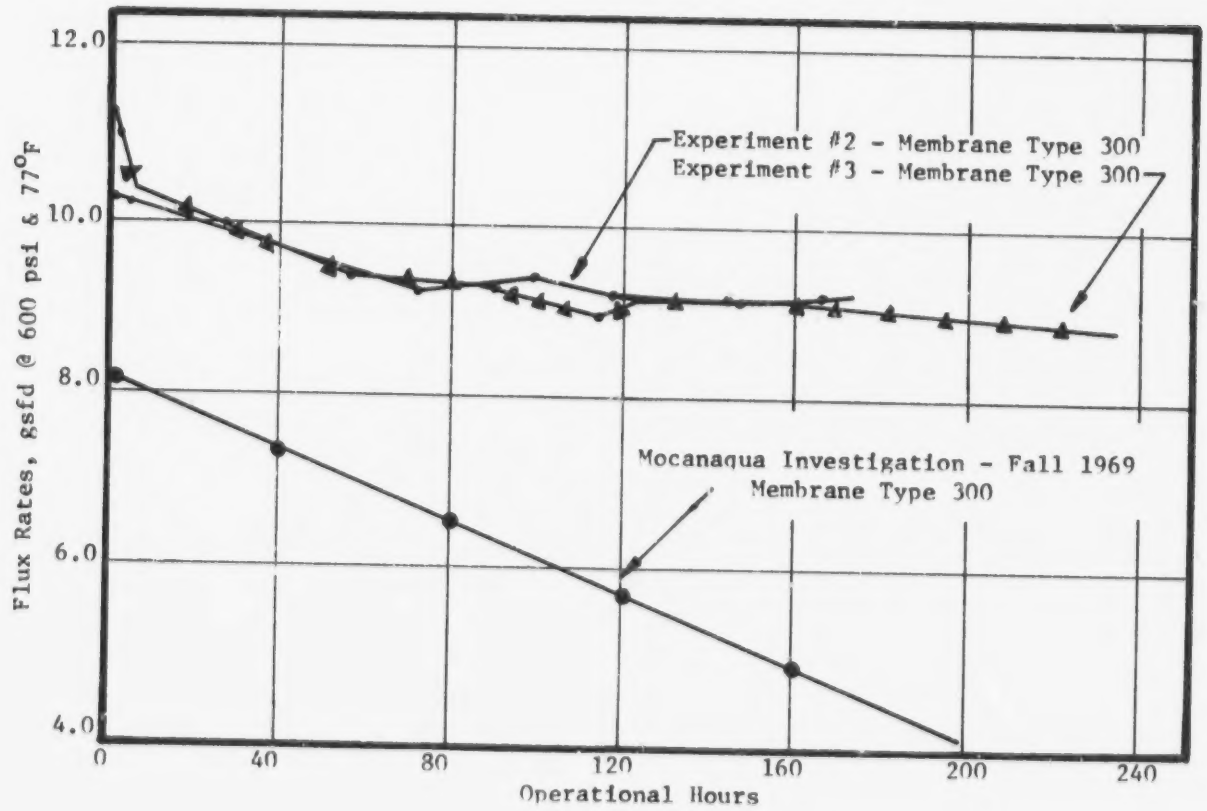

FIGURE 6

IRON FOULING INVESTIGATIONS (Type 300 Tubular Modules)

1. Only 13 gallons of actual acid mine water was mixed with 50 gallons of simulated acid mine water. Such a dilution might not have had a sufficient concentration of iron bacteria to cause significant membrane fouling during the short test period.

2. It was found that some copper was being dissolved into the aicd mine water from a bonze flow meter during the course of the experiment. Within six hours, a copper ion concentration of 12 mg/l was recorded in the feed water. Such a high level of copper ions could have had a toxic effect on the iron bacteria.

Although the cause of the iron fouling (previously encountered (3)) could not be pinpointed in the laboratory, it was concluded indirectly that biological iron oxidation was the cause of the observed membrane fouling. Such a conclusion was derived by the elimination of various other possible factors that could have caused the membrane fouling. For example,

1. Chemical oxidation of the iron did not occur as illustrated by the stable flux rates for experiments 1 and 2.

2. The new type 310 tubular modules did not have an effect on the flux rates, since the flux characteristics for both the type 300 and type 310 were similar (Experiments 1 and 2).

3. There were no apparent concentration polarization effects noted. with regard to chemical oxidation of the iron, as the brine velocity was varied (Experiment 1).

Investigation of Alternate Flow Schemes

Various items relevant to AMD treatment were evaluated during this phase of the laboratory studies. These included evaluating various types of RO hardware, as well as methods of pretreating the AMD prior to processing via RO. Also investigated were RO brine treatment alternatives as well as upgrading of the RO product water to meet USPHS potable water standards. As a result of these investigations, three possible AMD treatment schemes were considered. These schemes utilized similar supporting processes but each placed the RO unit at a different point within the scheme. This placement had a strong influence on the operation of the RO unit. Each of the three schemes was designed as a complete system to produce a potable water and to provide for ultimate disposal of all residues.

Treatment Method (A) - Limited Raw Water Pretreatment

Figure 7 presents one of the treatment schemes. This scheme provided filtration and bacteria control of the raw acid mine water prior to treatment by reverse osmosis. However, these pretreatment steps were not necessary for simulated acid mine water. The spiral wound and tubular RO units were operated under this scheme. It was not possible to operate the hollow fiber system on this feed water because the module had an aluminum shell and the effects of the low pH on the B-9 membrane

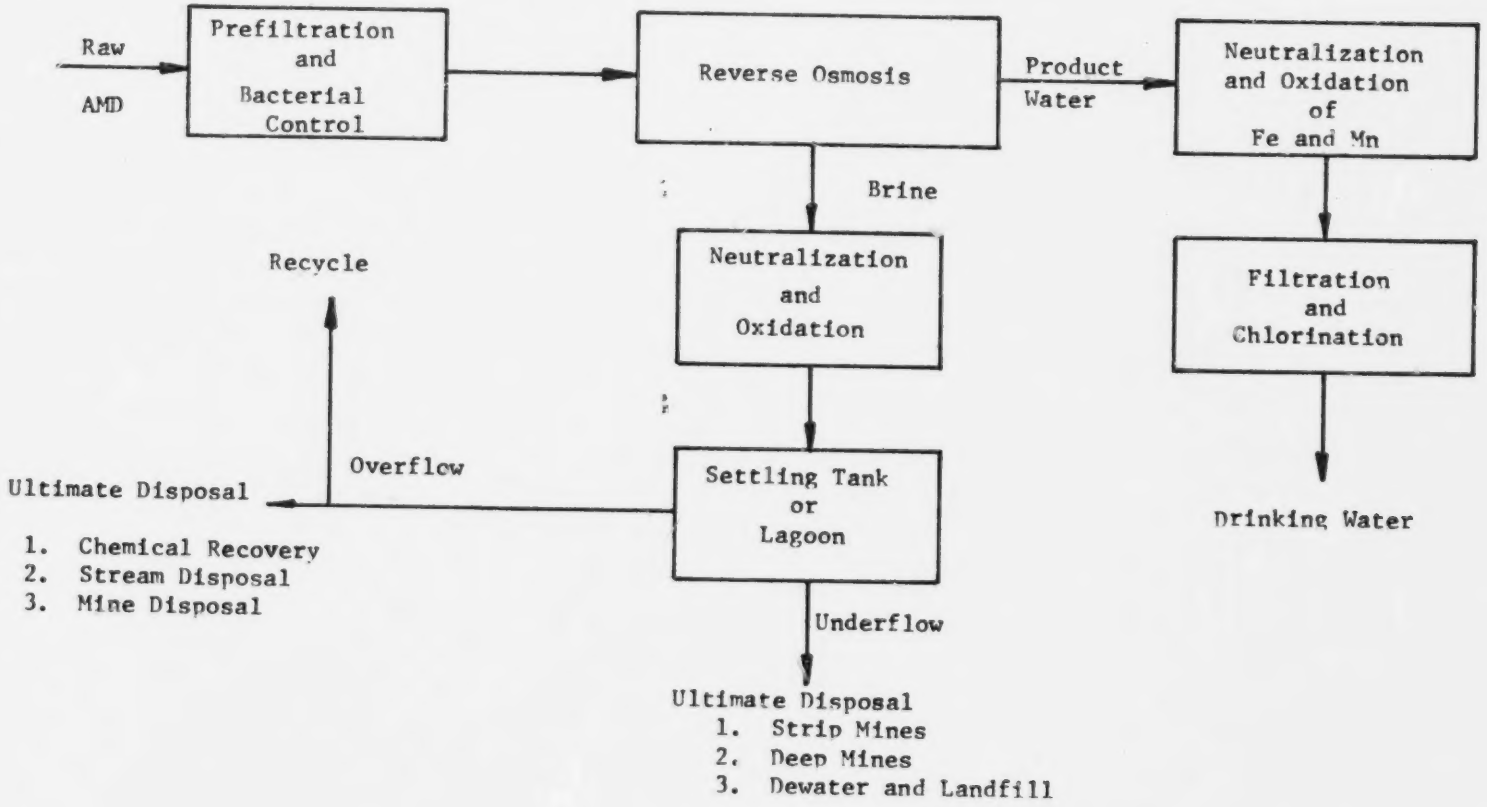

FIGURE 7

ALTERNATE AMD TREATMENT SCHEME - METHOD A, LIMITED PRETREATMENT

were not known at the time the testing was performed. (Note: later field operation indicated the membrane could withstand pH as low as 2.6 without damage.) A summary of the data taken on the spiral wound and tubular units is presented in Table 3. It was noted that the spiral wound membrane provided both superior rejection as well as flux. The greatest differences in rejections were in the iron and sulfate values. The large difference in flux can be partially attributed to the fact that the spiral wound unit had less than 10 hours of operation and the initial compaction of the membrane had not yet occurred. According to Gulf Environmental Systems, approximately 20% of the initial flux rate is generally lost in the first 100 hours of operation. This would yield a stabilized flux value of about 18-19 gallons per sq ft per day at 77°F compared to the tubular unit flux of 12 gallons per sq ft per day.

TABLE 3

DATA SUMMARY ON SIMULATED AMD

	Spiral Wound Unit		Tubular Unit	
	Feed Quality mg/l	Product Quality mg/l	Feed Quality mg/l	Product Quality mg/l
Iron	131	0.38	120	1.9
Calcium	190	2.0	134	2.2
Magnesium	106	1.0	98.5	1.9
Manganese	24	0.3	16.0	0.3
Sulfate	1250	3.0	1200	35
TDS	1643	26	1700	60
pH	3.5	--	3.6	3.4

NOTES:

Pressure	600 psig	600 psig
Feed Flow	5.06 gpm	1.15 gpm
Brine Flow	3.17 gpm	0.69 gpm
Membrane Area	150 sq ft	69.4 sq ft
Flux	24.4 gsfd @ 77°F	11.8 gsfd @ 77°F
Run	15.3 hours	325 hours
Recovery	37.3%	40%
Date Run	16 Dec. '70	September, 1970

The product water stream from both these units contained iron and manganese in excess of the USPHS drinking water standards, and the values recorded would be even higher if the product water recovery is increased. (This test was at 40% recovery.) The product water will therefore require post treatment to meet USPHS standards. Preliminary post treatment investigations of the RO permeate from the tubular system were made. The treatment included neutralization and oxidation followed by sand filtration. Two treatment methods were investigated. In Method I, the RO permeate was passed through a granular limestone bed for the neutralization and oxidation of iron (II), to iron (III) in the presence of oxygen. The detention time in the limestone bed was 3.8-7.5 minutes. The precipitated ferric hydroxide was removed by sand filtration. In Method II, the ferrous iron in the RO permeate was oxidized by the addition of 1.5 mg chlorine per mg ferrous iron after raising the pH to 7.0 by the addition of 25 mg/l lime. Oxidation time was 5 to 10 minutes. The resulting hydroxide was filtered through a sand bed. It was found that in both the schemes the sand filtered effluent contained less than 0.05 mg/l total iron. Additional data is required to define the process variables. However, it was shown that it is feasible to reduce the iron concentration well below the drinking water standards by providing any of the post treatment schemes described above. It will also probably be necessary to add a small amount of potassium permanganate to insure oxidation and precipitation of the manganese to the levels required by the USPHS standards. It may also be noted that conventional water treatment methods would be applicable for treating the permeate, since the levels of iron and manganese are similar to those found in many municipal water supplies.

The concentrate stream from RO is a potential pollution problem and further treatment is required. Typically this would involve neutralization and oxidation of the iron followed by solids/liquid separation in a settling lagoon. This treatment would remove essentially all of the iron and aluminum if operated at a pH higher than 7.5. Some $CaSO_4$ precipitation and removal can be expected, but the settling tank overflow would still be supersaturated with $CaSO_4$. The theoretical solubility of $CaSO_4$ is generally assumed to be 2000 mg/l but this can vary considerably with ionic strength of the solution, temperature, reaction kinetics and the concentrations of various other ions in the solution. It has been reported (8) that $CaSO_4$ precipitation within an RO unit can be controlled if the $CaSO_4$ concentration does not exceed about 300-400% of the theoretical concentration. The concentrate (brine), in this case, retains all the $CaSO_4$ in solution until the brine has passed out of the RO unit. Then if the brine is held for a period of time calcium sulfate will precipitate. The time dependency of $CaSO_4$ precipitation was also verified in this study. Simulated AMD was neutralized and settled to remove iron. This water was then passed through the hollow fiber RO module operating at 85% recovery. The resulting brine was supersaturated with $CaSO_4$. The changes in brine quality with time are shown in Table 4.

TABLE 4

CHANGE IN BRINE CHARACTERISTICS WITH TIME

Time (days)	Total Hardness (mg/l as $CaCO_3$)	Calcium as Ca	Sulfate (mg/l)
0	5800	1260	6000
1	4550	860	4500
4	3880	624	4200

It may be seen that after 4 days, the calcium concentration had been reduced to 624 mg/l which is extremely close to the theoretical level of 590 mg/l Ca which is equivalent to 2000 mg/l $CaSO_4$. These facts indicate that after treatment of the brine and precipitation of iron, aluminum, and $CaSO_4$, it may be possible to recycle the brine back through the RO unit, and hence eliminate a liquid waste stream. This system of recycling treated brine was developed at the EPA-Mine Drainages Field site and reported by Hill, et al. (9). The process (called Neutralosis) was operated for only short periods of time, and hence, additional data is required to fully evaluate this treatment method. It should also be noted that the Neutralosis process will not remove magnesium (Mg) or manganese (Mn) unless neutralization of the brine is taken to pH 9.5 or above. Therefore, these compounds would build up within the system until an equilibrium is reached where the pounds of Mg and Mn leaving the system in the waste sludge plus the product water would equal the pounds of Mg and Mn entering the system in the feed water. While no problem with magnesium was anticipated, since magnesium salts are quite soluble, the manganese could cause problems resulting in excessive amounts of manganese in the RO product water.

Because of the problem of various ion build-ups in the system when brine recirculation is practiced, experiments in brine softening, i.e. removal of Ca, Mg, and Mn were conducted. The brine produced with the hollow fiber unit from preneutralized and settled AMD has a composition as shown in Table 5. This brine was then subjected to various lime and soda ash dosages to determine how much chemical was needed to achieve a given degree of softening. Lime was added first at dosages of 0 to 2200 mg/l and reacted on a Phipps-Bird stirrer for forty minutes to provide contact between the sludge blanket and the brine. After settling, the supernatant was analyzed for total hardness, calcium hardness, and pH. In some cases sulfate analysis were also performed. The results of these tests are presented in Table 6.

TABLE 5

CHEMICAL ANALYSIS OF HOLLOW FIBER RO BRINE (1)

Constituent	Concentration- mg/l
Sulfate	6000
Total Hardness (2)	5800
Calcium Hardness (2)	3150
Manganese	65
Total Iron	0.35
Total Dissolved Solids	8085

(1) Raw AMD was preneutralized and settled
(2) As $CaCO_3$

TABLE 6

ANALYSIS OF SOFTENED RO BRINE

Lime Dosages (mg/l)	Soda Ash Dosage (mg/l)	pH	Total Hardness* (mg/l as $CaCO_3$)	Calcium Hardness* (mg/l as $CaCO_3$)
1500	0	11.6	5450	5450
1500	5000	11.6	450	80
1500	3500	--	1160	1040
0	4000	10.5	620	220

* As $CaCO_3$

It is apparent from Table 6 that 1500 mg/l lime (as CaO) removed all the magnesium hardness as well as all manganese but increased the calcium hardness. It may also be noted that 4000 mg/l soda ash was almost as effective in softening as 1500 mg/l lime and 5000 mg/l soda ash. From this it can be concluded that, it is not necessary to add lime to remove the magnesium hardness, since $MgCO_3$ is being precipitated. The addition of lime to precipitate magnesium hardness only increases the calcium noncarbonate hardness, thus requiring a larger soda ash dosage.

It is difficult to compare these laboratory dosages with the theoretical amount required due to the changing nature of the brine as discussed previously. Obviously, calcium sulfate precipitated with brine aging (Table 4), rendering it difficult to compare some of the tests. No calcium sulfate was precipitated during the softening

reactions, however, as the sulfate concentration did not change. The insensitivity of the sulfate test (±10%) make accounting for all the sulfate difficult. It is apparent from these tests that 4000-5000 mg/l soda ash will remove up to 90% of the total hardness present in the brine. It may not, however, be necessary to remove 90% of the Ca, Mg, and Mn to successfully recycle the supernatant back to the RO system. There are additional problems which must be evaluated regarding softening. These include increased sludge volumes which complicates the ultimate disposal problem, and methods of controlling the system.

It was therefore concluded that traatment method A was feasible and the question of recycling clarifier overflow requires additional field operation data. It is also recommended that additional studies be undertaken regarding brine softening.

Treatment Method B - Preneutralization (Figure 8)

In this scheme, the AMD was neutralized to a pH of 6.7 to 7.0 and aerated to provide oxidation of the iron and manganese. The objective was to reduce the soluble iron and manganese in the pretreatment step, to such a level that USPHS standards could be met after the reverse osmosis treatment. The resultant slurry would contain all the iron and manganese in an insoluble colloidal state which would be rejected 100% by the RO unit. All other soluble ions (Ca, Mg, and SO_4) would be rejected to the same extent as in untreated AMD. The product water from the RO unit would meet USPHS standards and following chlorination could be used as a potable water supply without post treatment as in the previously discussed treatment method. The brine stream from this system could then be routed to a lagoon for further concentration. Disposal of the overflow and underflow from the lagoon would be identical to the previously discussed treatment method (Method A, Figure 7). To evaluate this scheme the simulated acid mine feed water was neutralized with 155 mg/l of lime (as CaO) to a pH of 6.7 and was aerated continuously to keep the ferric hydroxide slurry in suspension. The ferrous iron content of this slurry was less than 0.05 mg/l and total iron content was about 125 mg/l. The iron slurry was then fed to the tubular RO unit, since a slurry of this kind can only be treated by a tubular RO system. Both the concentrate and the product water were recirculated back to the feed tank and the flux and the water quality data were monitored at various intervals. Figure 9 shows the variation of the flux rates with the operational time. The flux rate characteristics observed in this experiment were found to be very favorable. The flux rate dropped from 10.6 to 9.6 gallons per sq ft of membrane per day (gsfd) in the first six hours as would normally be expected due to initial compaction at 600 psi, but significantly higher flux rates were recorded at later time intervals. The flux rate increased to a value of 10.8 gsfd at the end of 24 hours of operation and then stabilized at a value of 11.2 gsfd for a continuous test duration of 235 hours. The reason for such flux characteristics could

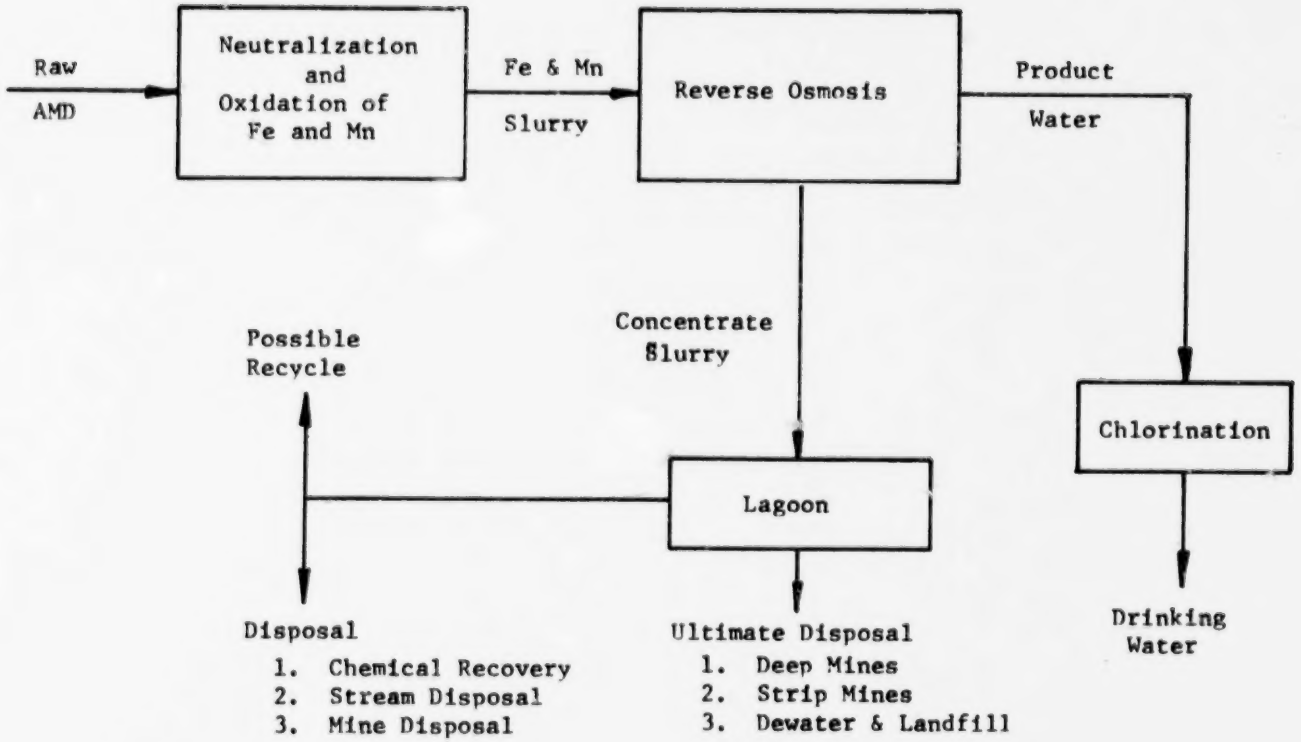

FIGURE 8

ALTERNATE AMD TREATMENT SCHEME - METHOD B, PRENEUTRALIZATION

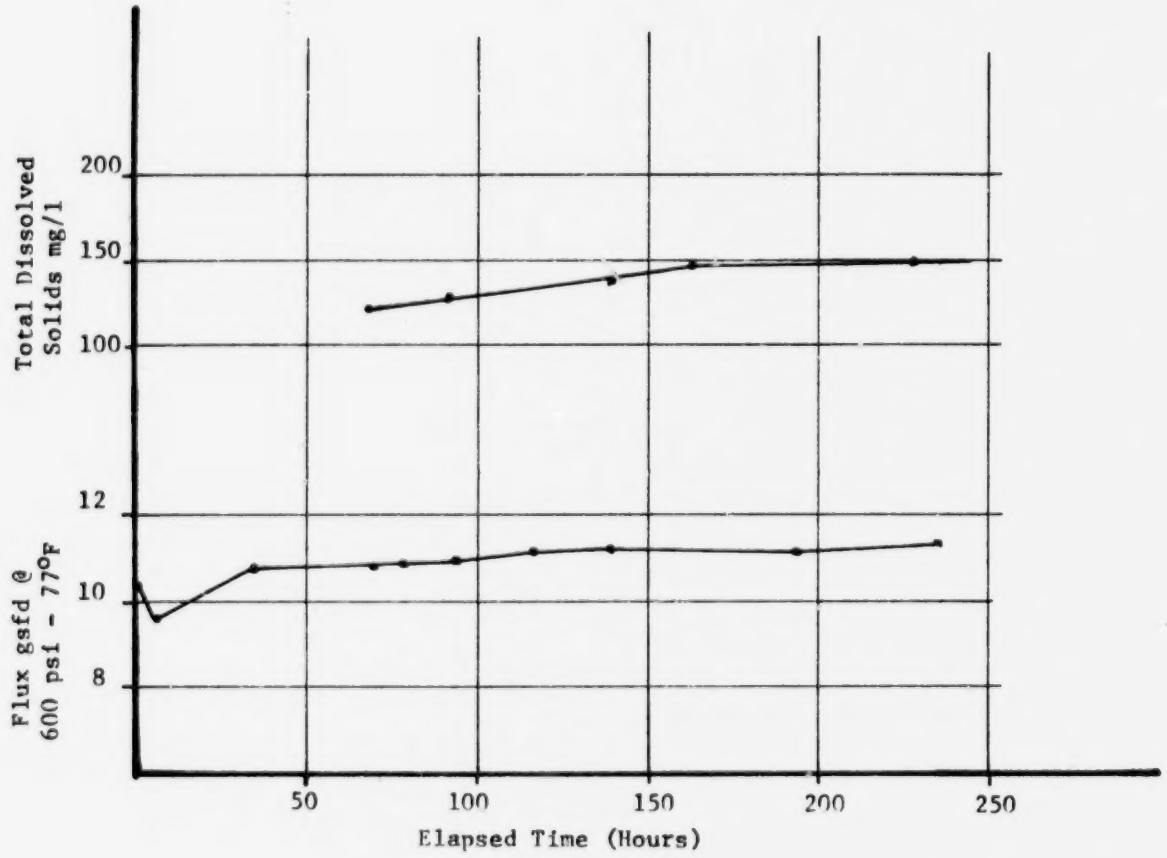

FIGURE 9

OPERATIONAL DATA FOR PRENEUTRALIZATION (METHOD B)

possibly be attributed to a scouring or brushing effect of the ferric hydroxide precipitate on the cellulose acetate membranes. This scouring apparently provided a continuous cleaning of the membranes and produced higher sustainable flux rates. Moreover, all the ferrous iron had already been converted to the ferric hydroxide precipitate and therefore no fouling of membranes because of the in situ oxidation of the ferrous iron was possible. During the experiment a slight increase in salt passage through the membrane was observed. This increase is also shown in Figure 9. It may be seen that the product water TDS increased steadily from 68 hours to 163 hours and then appeared to stabilize. This increase could indicate that some damage to the membrane desalting layer had occurred, and this could have been the cause of the observed flux increase.

Table 7 presents the typical feed and product water data obtained during this test. Also shown in Table 7 is the calculated product water at a 91% feed water recovery. It may be seen that even at high recovery (91%) the product water contained only 0.12 mg/l of iron. However, it may be pointed out that although the calculated total dissolved solids and iron concentrations at high recoveries were estimated to be well below the USPHS drinking water standards, a significantly high amount of manganese permeated the membranes. Table 7 indicates manganese concentration of 2.3 mg/l at 21% recovery and 4.8 mg/l at 91% recovery. The manganese in the feed waters was present in the manganous state; its oxidation to the manganic state is extremely slow below a pH of 9.0. Therefore, even when all the iron (II) had been converted to iron (III) at a pH of 6.7, most of the manganese was in the soluble manganesus state only, and hence, poorer removals were recorded for manganese ion by the RO membranes. The allowable concentration of manganese in drinking water is only 0.05 mg/l (USPHS standards). This means that even with a 99% rejection of the manganese ion by the membranes, its concentration in the permeate would exceed the drinking water standard limits for any feed waters containing more than 5 mg/l manganese.

TABLE 7
TYPICAL WATER QUALITY DATA FOR THE IRON SLURRY EXPERIMENT

Analysis	Water Quality		
	Feed	Feed Water Recovery	
		21%	91%*
pH, units	6.7	6.2	--
Total Solids	2100	140	305
Total Hardness as $CaCO_3$	1200	85	197
Calcium	280	23	49
Magnesium	120	8.5	18
Manganese	22.5	2.3	4.8
Sulfate	1300	99	213
Total iron	125	0.05	0.12
Ferrous iron	<0.05	<0.05	<0.10

NOTES : All quantities expressed in mg/l except where noted.
Membrane Type 300 tubular. *Calculated values

Additional laboratory tests were performed to investigate further removal of manganese in the pretreatment step, prior to RO. A sample of simulated AMD was neutralized to pH 7.4 with 170 mg/l lime as CaO. After aeration for 30 minutes and settling, the AMD contained less than 0.05 mg/l total iron. Manganese was also reduced to 14 mg/l. It was therefore demonstrated that although effective iron removal could be obtained by this treatment method, the removal of manganese was not very effective at neutral pH values. It was also determined that the oxidation of manganese (II) was extremely slow without the aid of an oxidizing agent other than molecular oxygen. The reduction of manganese to the level of about 5 mg/l is necessary to obtain a product water from RO meeting the USPHS Standard of 0.05 mg/l Mn. This is assuming a 97-99% rejection of manganese by the RO membrane. This level of rejection is possible with existing RO membranes. In an effort to reduce the manganese level to the desired value of 5 mg/l and also keep the pH below seven (the upper limit of cellulose acetate membranes), a series of chlorine oxidation tests were performed. Samples of previously neutralized and settled AMD were treated with varying amounts of calcium hypochlorite. The samples were then mixed for two hours, filtered and analyzed for residual manganese. The results are shown in Table 8. It can be seen that a large dosage of hypochlorite was needed to oxidize most of the manganese in two hours. However, the high chlorine residual remaining makes this type treatment undesirable.

TABLE 8

MANGANESE REMOVAL WITHOUT pH ADJUSTMENT

Hypochlorite Dosage (mg/l)	Residual Manganese (mg/l)	Final pH Units
5	10	6.7
10	9.5	6.3
15	7.6	5.8
20	6.5	5.5
44	1.0	5.6

Additional tests using potassium permanganate demonstrated that a dosage of 10 mg/l followed by 1/2 hour aeration at pH 7.0 would reduce the manganese to less than 1 mg/l. It was also found that neutralization to a pH of 9.5 would also reduce the manganese to about 1 mg/l. The permanganate method is preferred to the lime method, since pH 9.5 is above the working range of the RO membranes and would require additional pH adjustment, and addition of excess lime would present potential $CaSO_4$ scaling problems in the RO unit.

Another limitation on the treatment of acid mine drainage by this process scheme may be the attainment of higher feed water recoveries. It was found that when the iron slurry feed was concentrated fourfold by recycling the brine and wasting the permeate, calcium sulfate precipitate was observed in the brine. It is indicated that the maximum permissible feed recoveries without the calcium sulfate precipitation might be limited in the range of 75-80% in actual full size plant operation.

It was concluded that this scheme was not feasible based on the fact that possible membrane damage occurred and potential $CaSO_4$ scaling problems were produced.

Treatment Method C - Preneutralization and Settling

The third and final flow scheme preoxidizes the iron and manganese as discussed in Method B, but also provides for the removal of the precipitated compounds in a settling basin. This scheme is presented in Figure 10. The overflow from the settling basin is then processed by the RO unit. As in the previous method, the RO product waters would be suitable for potable use after chlorination. The concentrate from the RO unit may then be recycled to a limited extent as discussed for Method A. Table 9 presents a summary of the data taken on neutralized and settled AMD. All three units were operated on this feed water. It may be seen that the hollow fiber and spiral wound units gave comparable product water quality. The tubular unit gave considerably poorer water quality due to previous damage to these membranes while operating on an industrial waste. The expected water quality from the tubular unit would be very close to the tubular data shown in Table 3 if undamaged membranes had been utilized. If potassium permanganate were utilized during neutralization the product waters from this treatment method would meet USPHS standards.

Treatment of the brine would not be required as in Method "A", Figure 7, since the brine would already be neutralized. Recycling of any liquid streams back through the RO to reduce the volume of liquid for ultimate disposal would require the same consideration as disucssed for Method A. Method "C", Figure 10, also presents a potential $CaSO_4$ scaling problem, if lime is used to neutralize. This has to be considered a disadvantage for this method.

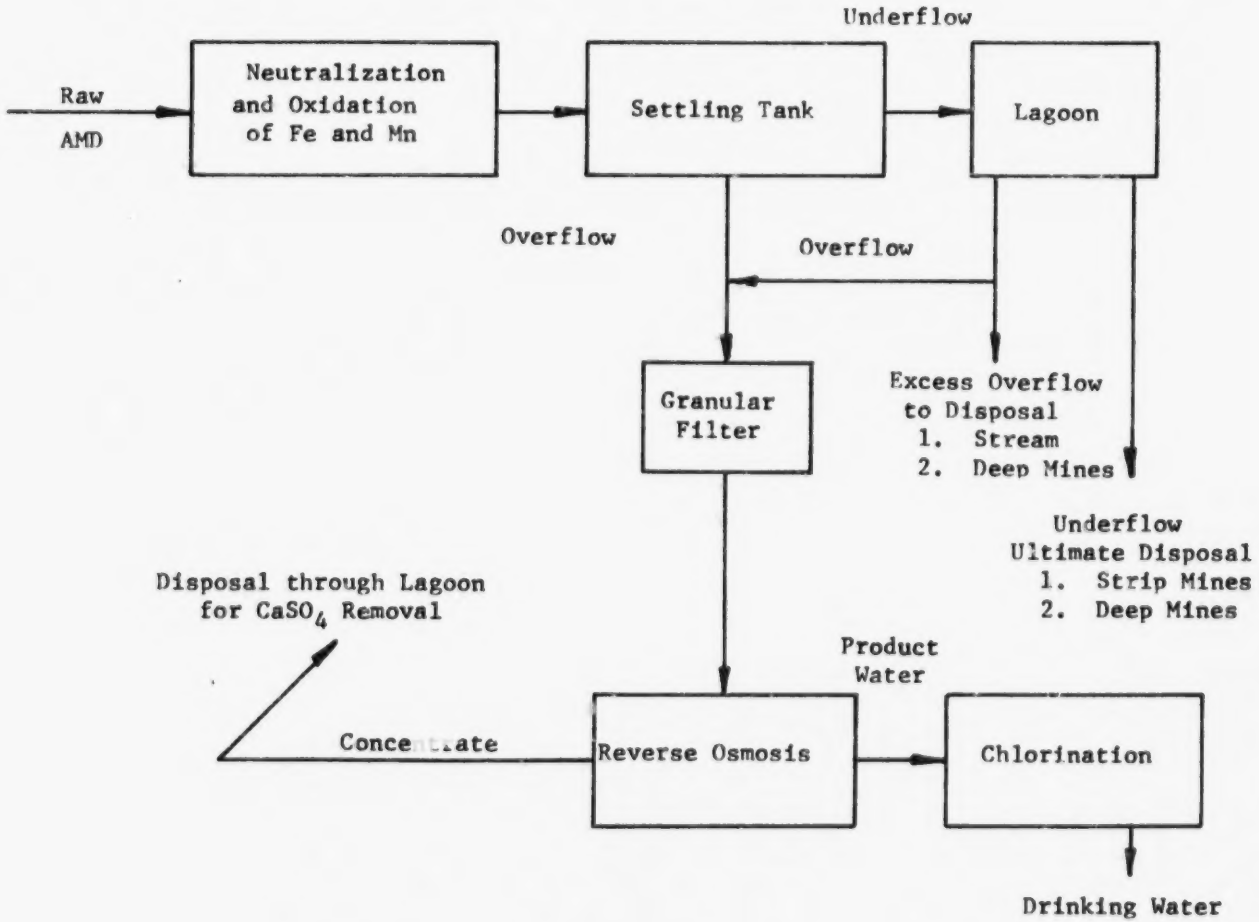

FIGURE 10

ALTERNATE AMD TREATMENT SCHEME - METHOD C, PRENEUTRALIZATION & SETTLING

TABLE 9

OPERATIONAL DATA FOR METHOD C
(FIGURE 10)

	Spiral Wound Unit		Tubular Unit		Hollow Fiber Unit	
	Feed Quality mg/l	Product Quality mg/l	Feed Quality mg/l	Product Quality mg/l	Feed Quality mg/l	Product Quality mg/l
Iron	0.2	0.05	0.2	0.05	0.2	0.02
Calcium	272	4.0	280	20	280	4.4
Magnesium	106	1.0	116	9.2	106	2.0
Manganese	14	0.2	14	1.4	13.6	0.2
Sulfate	1200	8	1250	88	1300	20
TDS	1691	29	1709	153	1701	25
pH	6.7	--	6.7	--	6.7	--

NOTES:

Pressure	600 psig	600 psig	400 psig
Feed Flow	4.62 gpm	1.08 gpm	2.65 gpm
Brine Flow	2.76 gpm	0.71 gpm	1.55 gpm
Membrane Area	150 sq ft	42 sq ft	1900 sq ft
Flux	22.8 @ 77°F	11.7 gsfd @ 77°F	1.2 gsfd @ 77°F
Run Length	3.0 hours	4.0 hours	4.5 hours
Recovery	40.4%	34.4%	41%
Date	16 Dec. '70	17 Dec. '70	18 Dec. '70

Note: Simulated AMD was neutralized and settled

SECTION VI

DESCRIPTION OF FIELD DEMONSTRATION SYSTEM

The laboratory investigation phase provided the necessary process information which was the basis for the field evaluation system design. The most significant conclusion from the laboratory work was that the cause of the previously encountered iron fouling was a result of bacterial oxidation of ferrous iron and subsequent precipitation on the membrane surface. It was theorized that to eliminate this source of fouling disinfection of the AMD was required. After considering various process problems based on the laboratory work and the work of others (4), a flow sheet was selected. A schematic diagram of the flow sheet may be seen in Figure 11. A photograph of the system is shown in Figure 12.

The feed water was pumped into the treatment system utilizing a PVC centrifugal pump (Item 1, Figure 12). This pump had a capacity of 15 gpm at 30 psi discharge pressure. This pressure was sufficient to route the feed water through the pretreatment system. This system consisted of a pressure sand filter which utilized 18 inches of filter sand with an effective size of 0.45 to 0.55 mm supported on 8 inches of gravel plus 1/4 inch, minus 5/16 inch size. From the sand filter the water flowed through two standard 9¾" x 2½" cartridge filters in parallel (Item 2, Figure 12). The function of the dual filtration system was to remove any suspended material, including bacteria, which might foul the RO membranes. The feed AMD after filtration was passed through an ultraviolet light sterilizer (Item 3, Figure 12). The sterilizer was a standard model 1000-S manufactured by Ultradynamics Corp. This unit provided an excess of 30,000 micro-watt seconds per sq cm of 2537 angstrom ultraviolet light and meets all U.S. Department of Health requirements for UV light purification equipment. The unit was equipped with two 15 watt UV bulbs. The volume of the radiation chamber was 8.05 gallons. Maximum depth of the radiated liquid was three inches. The unit was constructed of stainless steel. After the pretreatment system the flow was pressurized utilizing a moyno screw pump. This pump provided pressurized water to the two RO systems, i.e. tubular and hollow fine fiber.

The tubular system was manufactured by Calgon-Havens and utilized type 310 integral head modules. The basic system is the same one utilized in a previous study (3). The module arrangement, however, was modified for this study. The module arrangement utilized may be seen in Figure 13. It is basically a 6 x 4 x 2 array. Each row contained 5 modules in series. Each module has 16.9 sq ft of membrane area. The last two modules in each row of bank 2 (Figure 13) and the last three modules in each row of bank 3, contained volume displacement (turbulence promoter) rods (VDR). The purpose of these rods was to insure turbulent conditions even through the brine flow rate was being reduced. Turbulent conditions are desirable to prevent concentration

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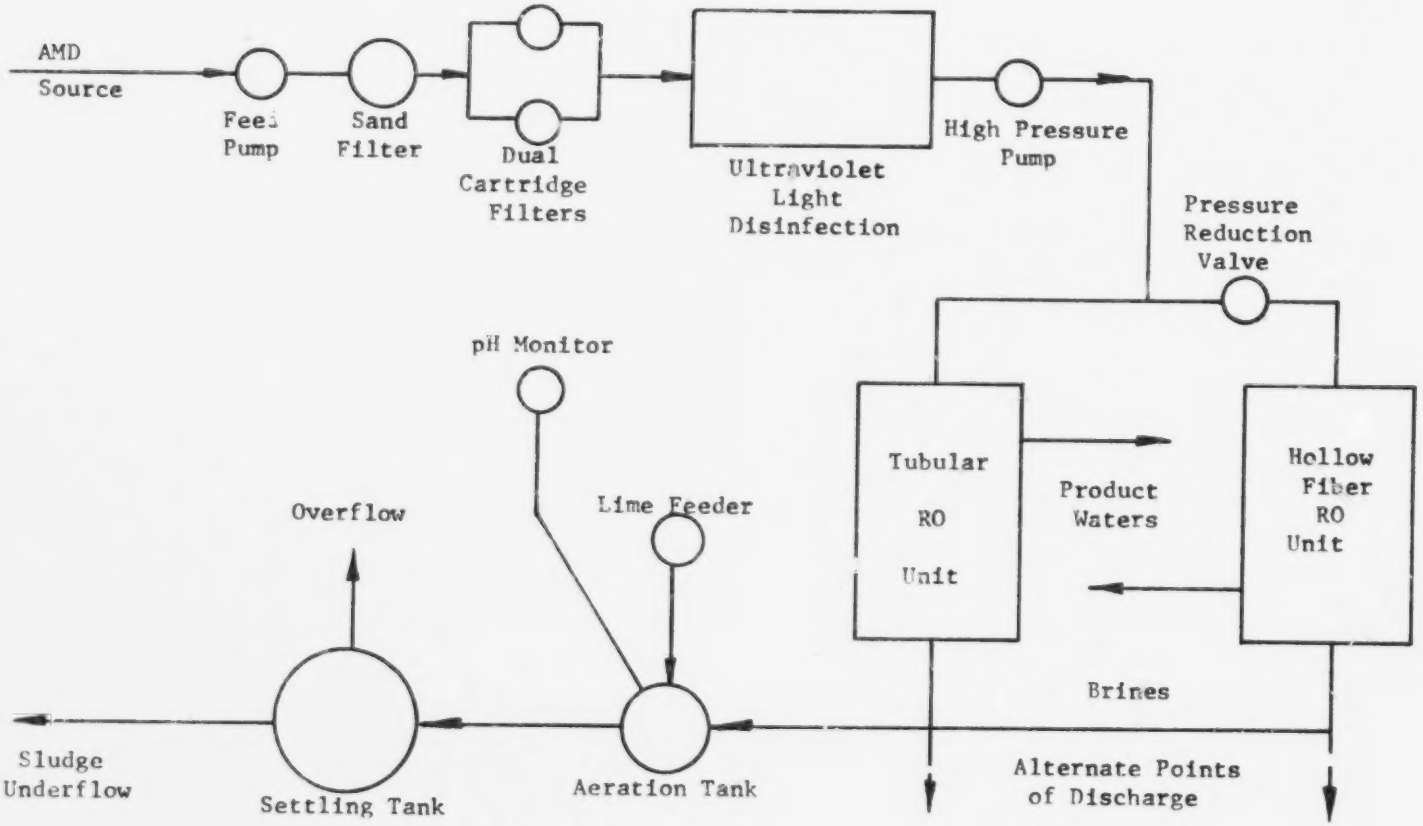
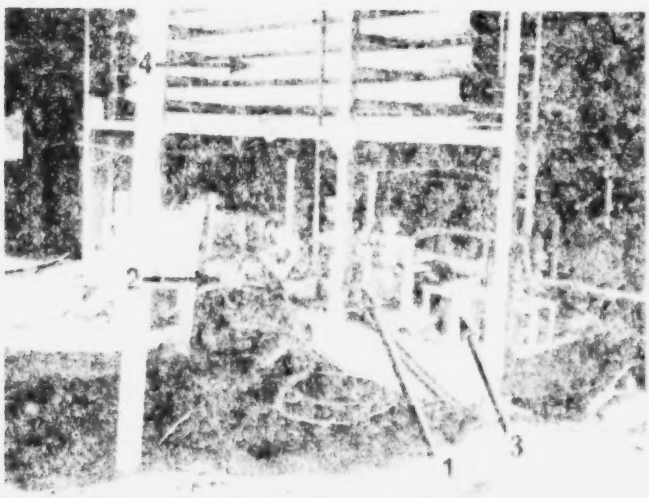


FIGURE 11

SCHEMATIC DIAGRAM OF THE FIELD TEST APPARATUS

a. Overall Arrangement



b. Pretreatment

1. Feed Pump
2. UV light
3. Cartridge Filters
4. Tubular modules

FIGURE 22

FIELD TEST APPARATUS

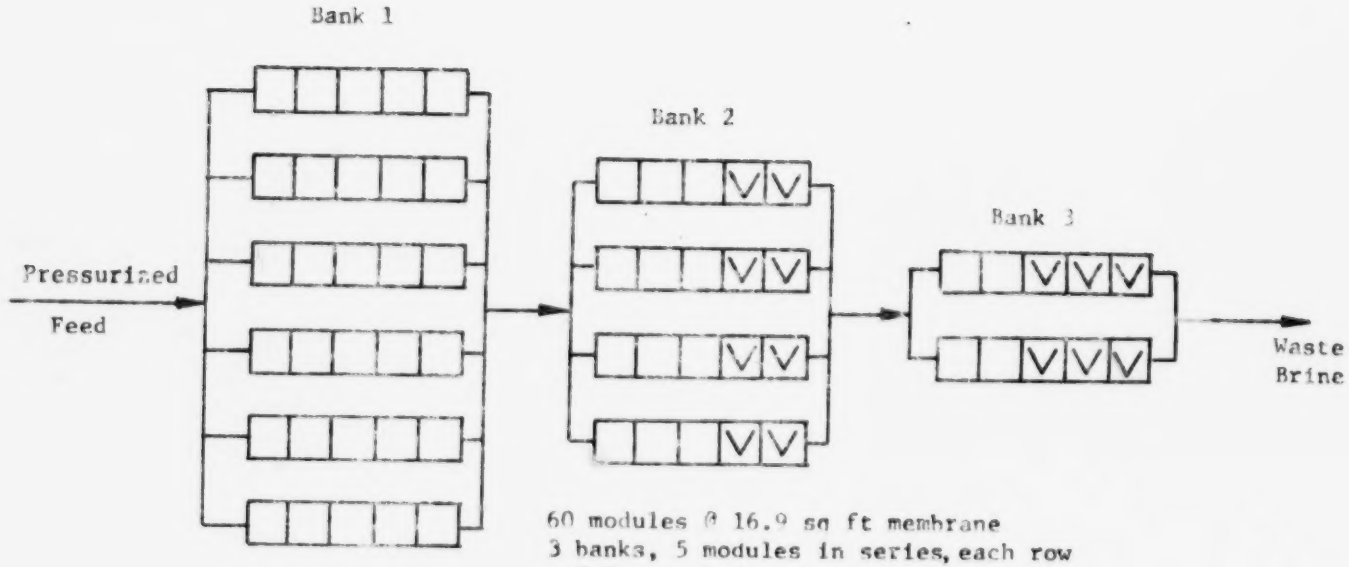


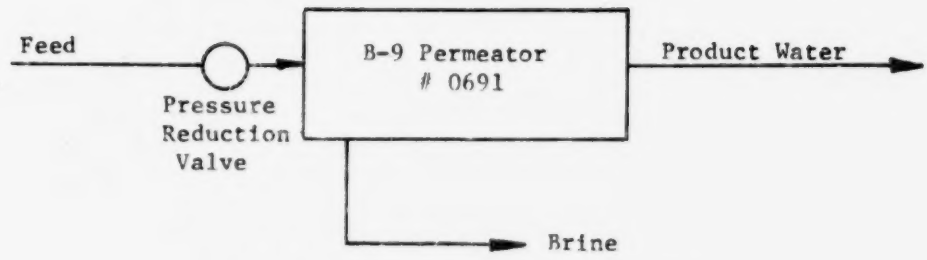
FIGURE 13

INITIAL MODULE ARRANGEMENT TUBULAR RO SYSTEM

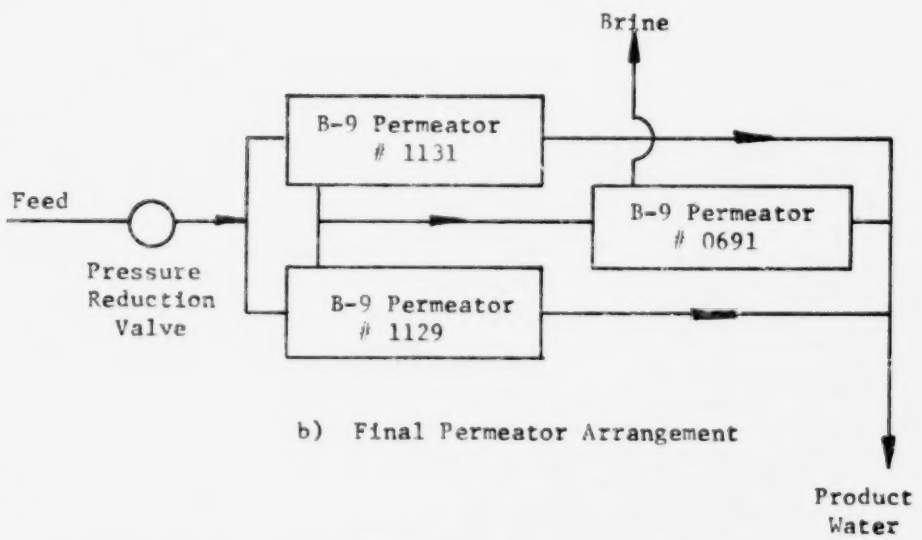
polarization induced fouling of the RO membranes. At times during the study the module configuration was changed to study various parameters. These changes generally consisted of altering the module configuration as well as the location of the (VDR) modules within the system. These changes and results of these changes are discussed in Section VII. The tubular modules are shown in Figure 12, Item 4.

The hollow fiber system which was utilized was manufactured by E.I. DuPont. The module arrangement for this system may be seen in Figure 14 and a photograph in Figure 15. The DuPont module contains approximately 1900 sq ft of membrane area in the form of hollow fine fibers. The fibers have an 85μ outside diameter and a 42μ inside diameter. The fiber is designated as B-9 and is an aromatic polyamide polymer. It is an ansitropic membrane with a 0.1μ skin. The fibers are packaged in a module called a permeator which measures 5½ inches outside diameter and is 47 inches in length. The rated water capacity of these permeators is 2000 gpd at 68°F and 400 psig pressure. The initial configuration (Figure 14a) contained a single permeator. Later in the study two additional permeators were added and utilized in a 2-1 array as shown in Figure 14b. Details on the operation of the hollow fiber system may be found in Section VII of this report.

The field demonstration system also included a brine treatment unit (Figure 11). This consisted of neutralization and oxidation of the brine followed by sedimentation. Photographs of the neutralization system are shown in Figure 16. The aeration tank was 5½ feet diameter round polyethylene tank which was operated at a 20 inch water depth for a total volume of 296 gallons. Lime was added to the brine flow utilizing a dry lime screw type feeder. The feeder was controlled by a pH meter to maintain the desired pH level. After aeration and conversion of the ferrous iron to ferric iron, the slurry was routed to a settling tank for solids liquid separation. The settling tank was a portable swimming pool 10 feet in diameter and was operated at 24 inch water depth. Water volume was 1172 gallons. An inlet baffle was provided (Figure 17, Item 4) to dissipate the inlet velocity and prevent short circuiting. The operating depth was held constant regardless of flow rate by a float operated throttling valve (Figure 16, Item 5). Settled sludge was removed manually utilizing a swimming pool vacuum cleaner type device (Figure 16, Item 6). The neutralization system could be operated on either tubular or hollow fiber RO brine. Details on neutralization system operation can be found in Section VII of this report.



a) Initial Permeator Arrangement

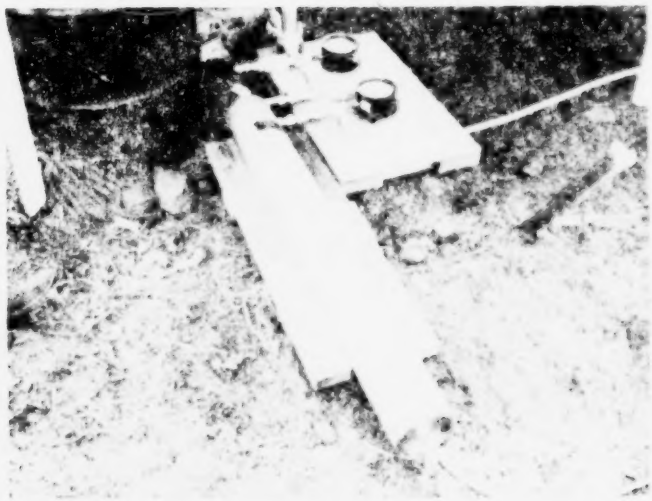


b) Final Permeator Arrangement

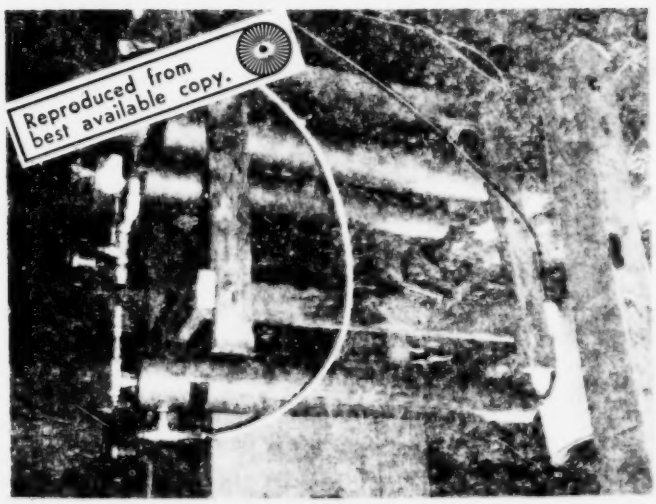
Each Permeator Rated @ 2000 gpd - 68°F - 400 psig
with 1900 sq ft Membrane

FIGURE 14

MODULE ARRANGEMENT HOLLOW FIBER RO SYSTEM



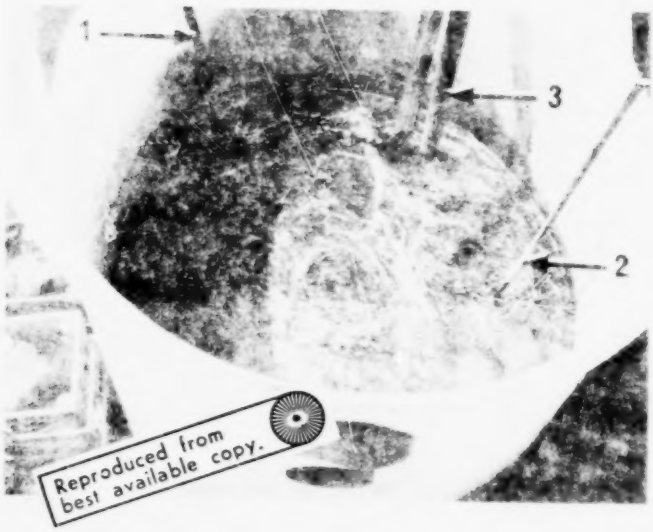
a. Initial System



b. Modified System

FIGURE 15

HOLLOW FIBER MODULES



a. Aeration Tank

b. Settling Tank

1. pH probe 2. Air Header 3. Neutralized brine inlet
4. Inlet Baffle 5. Effluent Float 6. Sludge Remover

FIGURE 16

NEUTRALIZATION SYSTEM

SECTION VII

DISCUSSION OF FIELD TEST RESULTS

The field operation covered many phases and utilized different RO equipment and configurations. A single hour clock was used as a reference for all field operations. This clock was not resettleable. At the start of this study the clock reading was 1140 hours. All figures and tables appearing in this section refer to elapsed operating hours which is the final clock reading minus the initial clock reading. In order to compare individual tables or figures with regard to absolute time a reference to the hour clock reading has been included in each table or figure where necessary. This allows comparison of tubular and hollow fiber RO system operation. The hour clock was connected directly to the high pressure pump and registered hours of pump operation. The field test phase started on April 30, 1971 and was completed on August 27, 1971. The total hours available during the test period was 2856. The elapsed operating time recorded was 2794 hours. The difference in these times was a result of power outages and system flushing.

Raw AMD Characteristics

The characteristics of the AMD utilized for this study are shown in Table 10. Analysis were run in the field using a Hach water analysis kit No. EL-DR. All analysis were run according to the instruments provided with this kit. However, all volumetric measurements for dilutions etc. were made using glass volumetric pipets and graduated cylinders instead of the less accurate plastic measuring devices provided with the kit. Measurements for pH were made with a Beckmann Model P-2 pH meter. Total dissolved solids (TDS) were measured with a Myron-L hand-held (TDS) meter.

Samples were also shipped to Milwaukee for laboratory analysis. Metal ion analysis was performed on a Perkin-Elmer Model 403 Atomic Absorption Unit. Laboratory TDS measurements were made using Standard Methods (10). Most samples were shipped to Milwaukee via airplane and the analysis completed within 24 to 30 hours after sampling. Some samples, however, were shipped parcel post. These samples were preserved with nitric acid as recommended by Perkin-Elmer.

Table 10 presents both the field and laboratory analysis, the mean values, and the 95% confidence range i.e. the range in which 95% of the analysis would be expected to fall. These analysis compare favorably with data taken from this site during a previous study (3) except for the iron values which were generally lower. All analysis were not run in the field, however Table 11 provides a direct comparison of

TABLE 10

RAW ACID MINE DRAINAGE CHARACTERISTICS

Home Office Laboratory Analysis[1]

Analysis	No. of Analysis	Mean Value	95% Confidence Range
Calcium mg/l	12	110.9	105 - 117
Magnesium mg/l	12	82.6	78 - 83
Manganese mg/l	12	14.0	13.5 - 14.5
Total Iron mg/l	12	70.1	64 - 76
Silica mg/l	11	10.8	10.3 - 11.3
Aluminum mg/l	12	8.3	8.0 - 8.6
TDS mg/l	11	1319	1234 - 1404

Field Analysis[2]

Analysis	No. of Analysis	Mean Value	95% Confidence Range
Calcium mg/l	34	114.6	109 - 120
Total Hardness mg/l $CaCO_3$	34	610	583 - 637
Total Iron mg/l	40	68	64 - 73
Ferrous Iron mg/l	40	63	59 - 66
Sulfate mg/l	37	798	757 - 839
TDS[3] mg/l	29	1185	1151 - 1219
pH units	23	3.38	3.27 - 3.43

1 Ion analysis by Atomic Absorption
2 Utilizing Hach Field Test Kid #CL-DR
3 With a "Myron-L" TDS meter

TABLE 11

COMPARISON OF FIELD[1] AND LABORATORY[2] ANALYSIS
OF RAW ACID MINE DRAINAGE

Elapsed Time Clock		Total Iron mg/l		Calcium mg/l		Total Hardness as $CaCO_3$ mg/l	
Field	Laboratory	Field	Laboratory	Field	Laboratory	Field	Laboratory[3]
1230	1234	78	71	130	115	648	835
1301	1302	74	74	130	115	728	884
1395	1380	71	75	152	115	600	851
1632	1590	54	54	96	97	480	707
1898	1899	55	57	122	97	465	710
2051	2051	56	59	120	102	550	746
2314	2314	57	69	116	107	560	780
2528	2528	70	74	120	112	650	808
3051	3074	76	84	124	125	700	927
3140	3162	83	80	100	123	690	895
Average		67.4	69.7	121	111	607	814
Range at 95% Confidence Level		±7.6	±7.1	±11	±7	± 65	±55

1 Using Hach Kit #EL-DR
2 Using Atomic Absorption
3 Calculated from Ca, Mg, Mn, Al, and Fe analysis

results from the laboratory and field. Comparisons were made for total iron, calcium and total hardness. Paired comparison tests were run (11) on each set of field and laboratory data. The results of this test indicated there was no significant difference between the field and laboratory analysis for iron and calcium at the 95% confidence level. Stated another way the apparent differences between the field and laboratory analysis for iron and calcium were not significant statistically. The conclusion can be made, therefore, that the field test procedures gave results for iron and calcium comparable to laboratory analysis.

The total hardness values for the field and laboratory however, were not consistent. The total hardness values for the laboratory analysis were calculated by summing the hardness producing ions (Fe, Al, Ca, Mg, Mn) and expressing this sum as total ($CaCO_3$) hardness. The laboratory total hardness values were always significantly higher than the field value. This indicated that the total hardness test in the field was not accurately measuring all the hardness producing ions. Furthermore, Standard Methods (10)states that the levels of certain ions (Fe, Mn) present in these samples will cause a low total hardness reading. Manganese and aluminum can also be analyzed with the Hach kit, however, they were not measured during this study. Based on previous experience, reasonable accuracy is expected on acid mine waters for these analysis. Since the total hardness value is used to calculate the magnesium level it must be concluded that magnesium cannot be accurately determined on acid mine waters using a Hach kit.

The dissolved oxygen levels in the AMD feed water were measured periodically. These values are shown below.

Dissolved Oxygen Levels in the AMD Feed Water

Elapsed Time Clock Hours	Dissolved Oxygen in Feed AMD mg/l
1189	0.6
1470	0.6
1565	0.9
1664	0.9

Note: All measurements made with a YSI (Yellow Springs) Dissolved Oxygen Meter

Since a relatively long suction line for the AMD feed water (120') was utilized, the dissolved oxygen level was checked at the AMD source and after the feed pump. These analysis indicated no air leaks were present in the suction piping.

Raw water quality variation is presented in Figure 17. A reduction in all values is evident at about a 1600 elapsed time clock reading. This was due to a large amount of rain which fell on the area during this period. This caused an increase in the water table and had the effect of diluting the feed water concentrations. This sharp drop in contaminent level was followed by a gradual increase until about 2900 hours when additional heavy rains were experienced which caused another drop in concentration levels.

Operation of the Pretreatment System

The pretreatment (prior to RO) consisted of filtration followed by ultraviolet light disinfection. Initially a sand filter followed by 5 micron cartridge filtration was utilized. The function of the sand filter was to protect the cartridge filters and hence provide longer runs. It was soon discovered that the sand filter was not removing any particulate matter and use of the sand filter was discontinued after about 100 hours of operation. The 5 micron polypropylene filters were manufactured by Pall-Trinity (Filter # MCY 1001 YCH2). They contained 3.7 sq ft of filter area. These filters (two in parallel) had a life of approximately 24 hours or 7,500 gallons per filter. Because of this relatively short life and the high costs involved, a switch was made to 10 micron filters. These were also manufactured by Pall-Trinity (Filter # MCY100 1EE) and contained 3.7 sq ft of filter surface. The 10 micron filters (two in parallel) had an average life of 108 hours which was equivalent to 32,400 gallons per filter. The great difference in filter life indicated a significant number of particles smaller than 10 micron and larger than 5 micron were present in the mine discharge. Since there was no apparent effect on the RO unit operation, 10 micron filters were used during the remainder of the study.

The purpose of the ultraviolet light disinfection unit was to kill the iron bacteria present in the mine drainage and, hence, prevent bacterial oxidation of iron II to iron III. Iron oxidation studies were performed to enable evaluation of the effectiveness of the UV light. Figure 18 presents the data collected on the iron oxidation rates of four different samples. The four samples were stored in polyethylene bottles. Once each day the cap was removed and then replaced to allow oxygen equilization and then shaken vigorously to insure an oxygen saturated sample. The raw AMD sample only filtered through a 10 micron filter exhibited a rapid and immediate reduction in iron (II) level. On the other hand a filtered sample in which the pH was adjusted to 2.5 experienced little iron (II) oxidation for a period of 100 hours and then a rapid reduction at approximately the same rate as the sample which had no pH adjustment. The sample which had been exposed to ultraviolet radiation had little decrease in iron II level until 100 hours and then a gradual decrease until no iron (II) remained at 350 hours. The sample disinfected with formaldehyde had little decrease in iron (II) level when the experiment ended at 676 hours (iron II was 42 mg/l at this time).

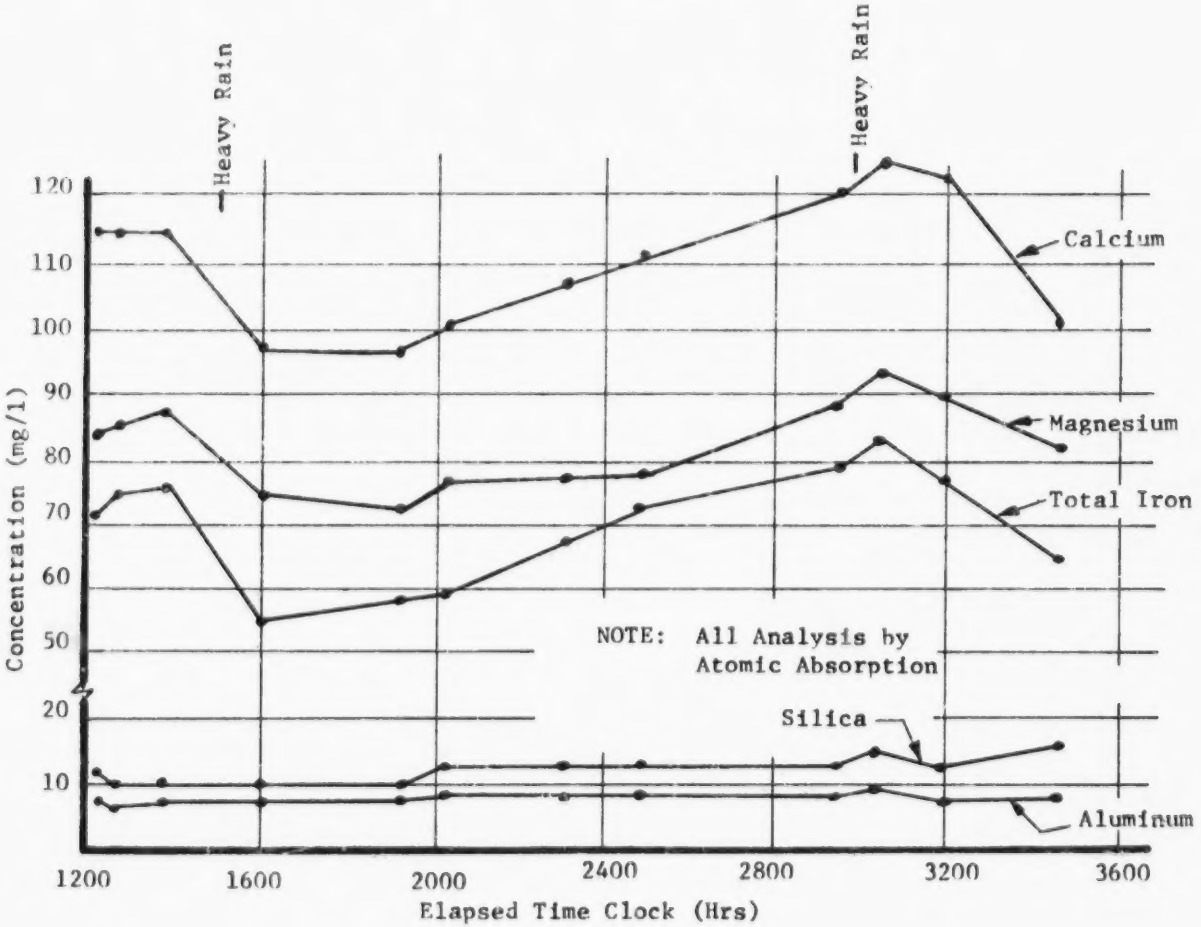
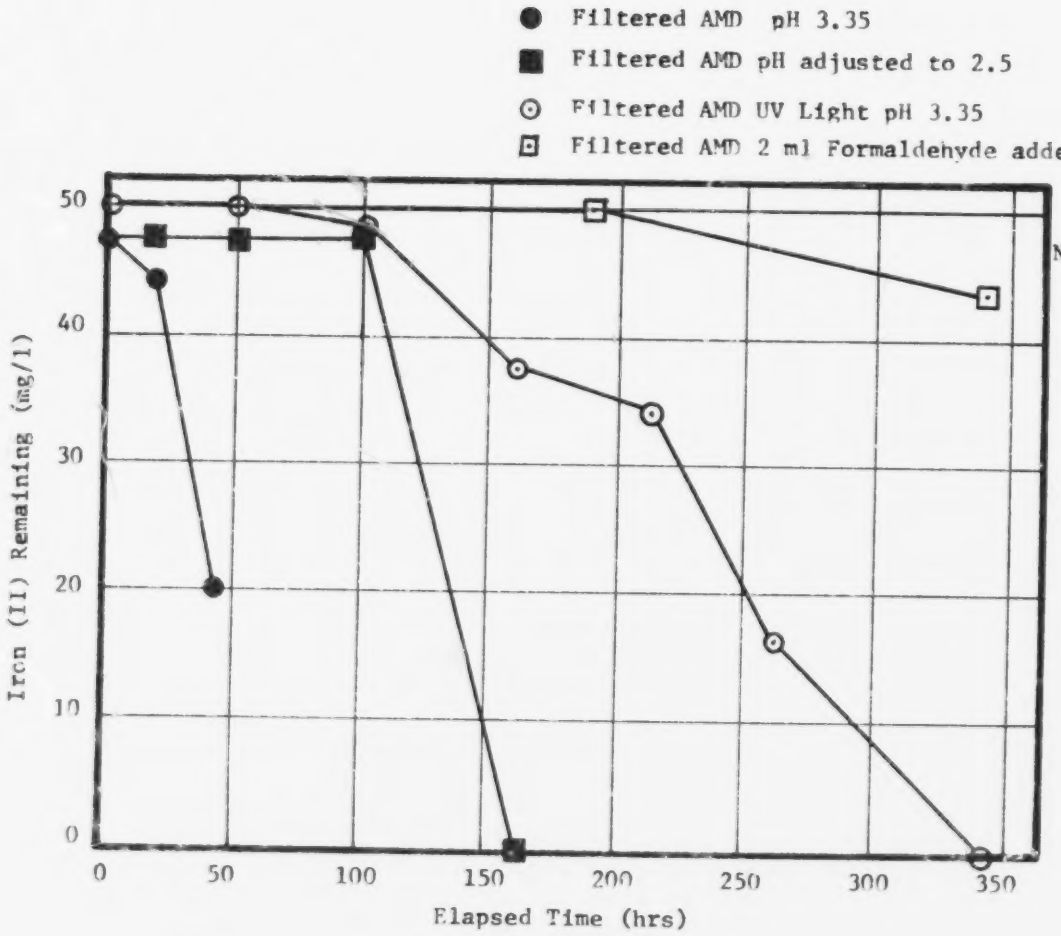


FIGURE 17

RAW WATER QUALITY VARIATION



NOTE: Total Iron = 54 mg/l for all samples. UV light radiation time = 53 seconds

FIGURE 18

IRON OXIDATION RATES

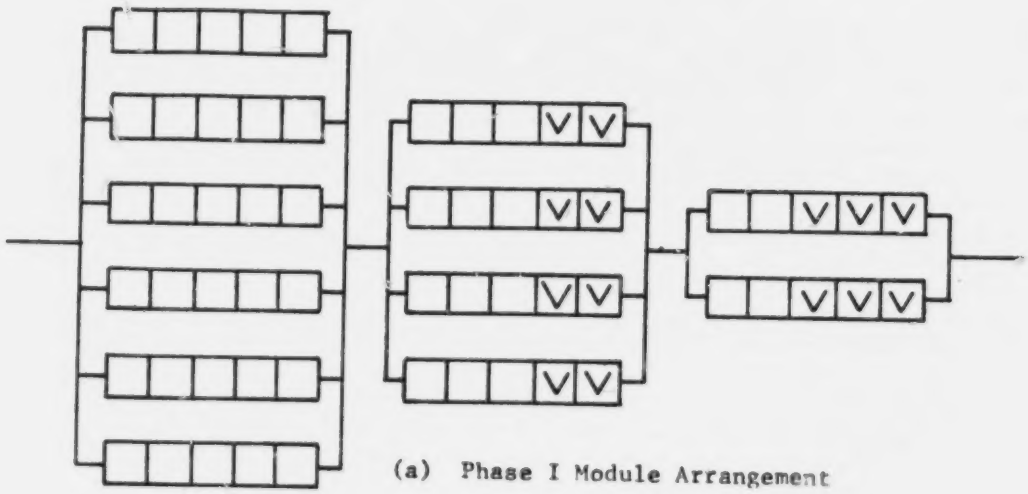
At the completion of this oxidation study it was obvious that iron oxidizing bacteria were responsible for the rapid decrease in iron (II) levels since the oxidation rate exceeded the natural chemical oxidation rate by many orders of magnitude (12). It was also apparent that the oxidation was inhibited by merely lowering the pH. This is not surprising, since this can upset biological oxidation until the organisms acclimate to the new pH level. Once this occurred the oxidation proceeded at the same rate experienced in the raw sample at pH 3.35. It may be noted (Figure 18) that the UV light arrested the oxidation for approximately 100 hours and a slower rate of oxidation persisted for the next 250 hours. It was apparent that the UV light did not effect a 100% kill, however the possibility of sample contamination during the frequent analysis does exist. The sample disinfected with formaldehyde was the most stable as demonstrated by the small decrease in iron (II) level. In any event addition of acid, UV light disinfection, or formaldehyde will provide protection against iron oxidation and subsequent fouling of the RO membrane.

Some mechanical difficulties were experienced with the UV light system. After 626 hours of use the bulbs (Model P-247) failed due to destruction of the end caps by the UV light. These were replaced with bulbs containing a foil wrap around the end caps. This wrap was apparently not effective since the bulbs again failed after only 740 hours of usage. At this point, a different model bulb was installed (P246). These bulbs lasted through the end of the study, an additional 900 hours without failure. Additional operating data is required to determine the life of this new model bulb. The manufacturer, however, guarantees the P-246 bulb for 7000 operating hours.

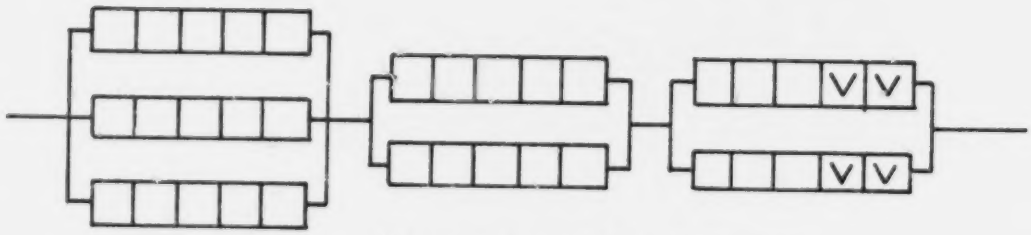
Operation of the Tubular RO System

The tubular RO system was operated in five separate phases during this study. Generally the phases were marked by different module configuratonns or new modules. Figure 19 presents the module arrangements utilized for the first four phases of the tubular study. Phase I utilized 60 Type 310 modules in a 6-4-2 array with 5 modules in series in each row. The last two modules in bank 2 and the last 3 modules in bank 3 (Figure 19a) contained volume displacement rods (VDR) which increased the brine velocity. The normal inside diameter in the tubular system is 1/2 inch which corresponds to a linear brine velocity of 1.64 fps per gpm of brine flow. The volume displacement rods effectively increase this velocity to 2.5 fps per gpm of brine flow. When utilizing VDR's the headloss through the module increases significantly. In a module without the VDR's the headloss at 1 gpm brine flow is 5 psi per module, while with VDR's this increases to 22 psi per module.

The Phase II study utilized 35 type 310 modules in a 3-2-2 array (Figure 19b). VDR's were utilized only in the last two modules of bank 3. Phase III and IV utilized 15 type 310 modules in series with



(a) Phase I Module Arrangement



(b) Phase II Module Arrangement

(c) Phase III & IV Module Arrangement

NOTE: Each square represents one module, modules marked V contain turbulence promoter rods.

FIGURE 19

MODULE ARRANGEMENTS UTILIZED FOR TUBULAR 310 MODULES

VDR's in the last two modules (Figure 19c).

The entire flux history for Phases I to IV is presented in Figure 20. During Phase I the product water flux decreased steadily from 13 gsfd to 8.5 gsfd in only 480 hours. It was originally assumed that the system had been contaminated with iron oxidizing bacteria because of trouble experienced with the UV light. At 160 hours the system was disinfected with a quatenary ammonium compound (L-11-X). This accounts for the slight increase in flux noted at 160 hours, since the membranes were relaxed (operated at low pressure) and this generally results in a flux increase for a short period of time. Immediately after the disinfection, the flux continued to decline rapidly. At this point it was felt that the brine velocities might possibly be too low and that concentration polarization effects were the cause of the rapid fouling. It was therefore decided to increase the brine velocities. This could not be accomplished with the configuration utilized in Phase I because of the high head losses experienced across the VDR modules. To reduce this head loss and increase the velocities the module arrangement was changed for Phase II (Figure 19b) to a 3-2-2 array. The minimum brine velocity was increased from 1.2 - 1.4 fps to 2.0 - 2.2 fps. Prior to starting Phase II operation, the modules were flushed with an ammoniated citric acid solution (1.5 wt % citric acid - buffered to pH 4 with ammonia) in an attempt to remove any iron fouling which may have occurred during Phase I. The system was then put into operation for Phase II. It may be seen in Figure 20 that the flux initially declined and then stabilized at a value of about 7 gsfd. The initial high flux readings were probably due to membrane relaxation, since it was later found the ammoniated citric acid was not effective in removing iron fouling from AMD fouled RO membranes. The stabilizing of the flux values during the latter part of Phase II could have been a result of the increasing brine velocities, but a definite conclusion cannot be made without additional study. It will also be noted in Figure 20 that the water recovery was lowered to about 50% in Phase II, and this may have had a stabilizing effect on the flux rates.

Phase III operation was identical to Phase II except the module configuration was changed to 15 modules in series (see Figure 19c). This change was made in order to reduce the high pressure pump requirements. As may be seen (Figure 20) the flux remained stable throughout the entire Phase III. The recovery was also increased to about 60% for the entire Phase III. At the end of Phase III, the back pressure valve clogged and the system ran for 10 to 12 hours with no brine flow. This resulted in the end modules becoming completely clogged with $CaSO_4$. It was therefore necessary to switch to 15 new modules and this marked the beginning of Phase IV as shown in Figure 20.

The initially high flux experienced in Phase IV was probably a result of relaxation (0 pressure) while the modules were not in use. The flux immediately began to decline to approximately the same levels experienced in Phase III.

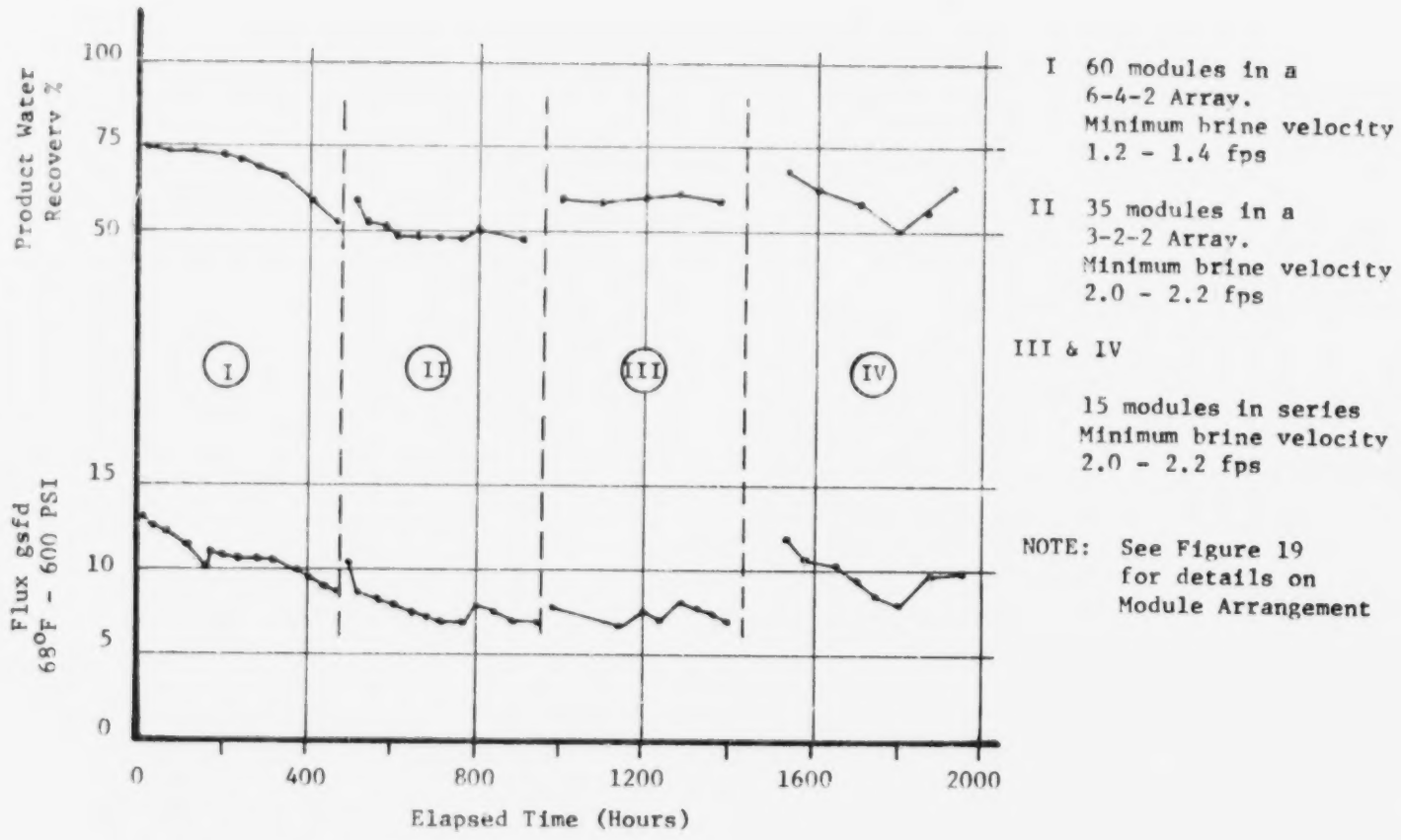


FIGURE 20

TUBULAR RO SYSTEM OPERATION WITH TYPE 310 MODULES

At the end of Phase IV it was noted that the flux declines experienced with the tubular system were not experienced with the hollow fine fiber system or a spiral wound system also operating at the same site (13). It was also noted that both the spiral wound system and the hollow fiber system had considerably higher salt rejection as compared to the tubular system. It was felt that this fact may have had some influence on the flux declines experienced. To test this theory, five high flux-high salt rejection modules (Type E610) were installed and put into operation. Figure 21 presents the flux history for these modules. An extremely high initial compaction set was experienced during the first 40 hours of operation. The flux then stabilized at about 15-16 gsfd until 240 hours elapsed time. A gradual decline then occurred through about 440 hours, at which time the flux stabilized at 12.5 gsfd for the remainder of the study. The flux decline experienced from 240-440 hrs was also experienced on the hollow fiber unit and was believed caused by a higher than normal iron (III) content in the AMD. The decline was entirely due to iron fouling, since operation at 40-45% recovery was well below the $CaSO_4$ fouling range. The modules were flushed with a sodium-hydro-sulfite solution (4 wt %) for one hour. This resulted in a dramatic increase in flux as shown in Figure 21. Since additional operating time was not available, it is not known how much of this flux increase was due to cleaning and how much was due to membrane relaxation. It is felt, however, that a substantial gain was accomplished, since membrane relaxation alone would not account for an increase in flux of about 33%.

The salt rejection properties for both the types 310 and 610 tubular modules are shown in Table 12. The salt rejection was calculated based on average brine concentration experienced on the membranes, i.e. the average of feed and brine concentrations. This procedure allows comparison of salt rejection, while operating the RO system at different product water recoveries.

Generally salt rejection for the type 310 modules was in the range of 98.5-99% for Ca, Mn, Fe, Al, and SO_4. Silica rejection was extremely low at an average of 46 percent. The type 610 modules had significantly higher salt rejection in the range of 99.5-99.6% for Ca, Mg, Mn, Fe, Al, and SO_4. Silica removal was also considerably higher than the 310 modules at an average of 93.9 percent rejection. No apparent changes occurred in the salt rejection throughout the operational period.

The mechanical operation of the tubular system was excellent. No module failures were experienced over the entire 2800 hour operation. This reflects the improvements made in tube construction since the last study (3). The problem experienced with the plugged back pressure valve at the end of Phase III was a result of the extremely low total brine flows during this time, and the fact that the high pressure pump was feeding two separate RO systems resulting in less than positive brine flow control. This problem is not anticipated in full scale systems.

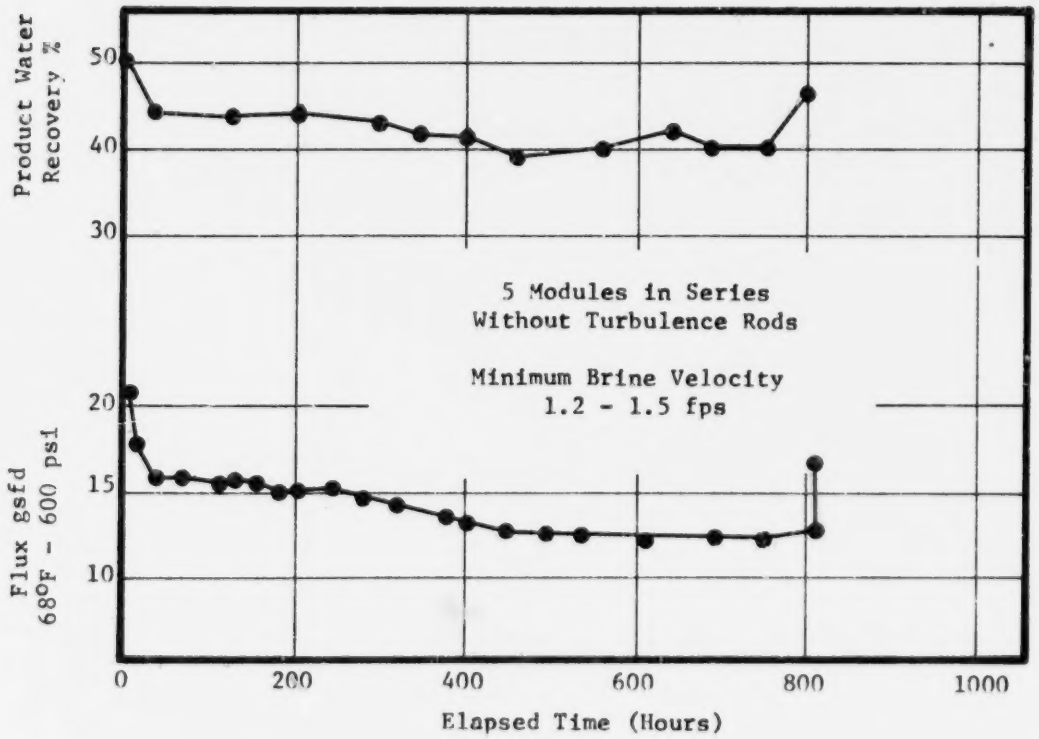


FIGURE 21

TUBULAR RO SYSTEM OPERATION WITH TYPE 610 MODULES

TABLE 12

SALT REJECTION CHARACTERISTICS
TUBULAR RO SYSTEM

Ion	Raw Water Quality mg/l	Type 310 Modules Brine Quality mg/l	Product Quality mg/l	Salt Rejection %	Type 610 Modules Brine Quality mg/l	Product Quality mg/l	Salt Rejection %
LABORATORY ANALYSIS							
Calcium	111 ± 6	287 ± 98	2.8 ± 0.8	98.55 ± 0.32	168	0.58	99.57
Magnesium	83 ± 4	220 ± 82	1.6 ± 0.4	98.89 ± 0.17	132	0.39	99.64
Manganese	14 ± 0.6	39 ± 11	0.31 ± 0.06	98.79 ± 0.19	22	0.08	99.55
Iron (Total)	70 ± 6	180 ± 71	1.42 ± 0.3	98.81 ± 0.21	102	0.31	99.55
Aluminum	8 ± 0.3	21 ± 6	0.2 ± 0.08	98.6 ± 0.53	14	0.1	99.61
Silica	11 ± 0.5	16 ± 2	7.2 ± 0.5	46.4 ± 5.0	22	1.0	93.9
Total Dissolved Solids	1319 ± 85	3523 ± 1100	53 ± 11	97.66 ± 9.82	2074	17	99.13
FIELD ANALYSIS							
Calcium	118 ± 6	373 ± 74	4.5 ± 1.1	98.24 ± 0.31	200	0.80	99.47
Total Hardness as ($CaCO_3$)	602 ± 42	1939 ± 378	22 ± 6	98.37 ± 0.37	1500	6.0	99.45
Iron Total	67 ± 5	226 ± 60	1.5 ± 0.2	98.94 ± 0.15	200	0.65	99.54
Iron (II)	64 ± 4	186 ± 37	1.4 ± 0.2	98.85 ± 0.09	160	0.62	99.49
Sulfate	774 ± 45	2056 ± 326	19 ± 4	98.73 ± 0.27	1450	2.0	99.80
pH	3.38 ± 0.06	2.96 ± 0.1	4.1 ± 0.2	--	3.0	4.0	--

Notes: All ranges shown at 95% confidence level.
Salt Passage (%) = 200 (Product Water Quality) / (Feed Quality + Brine Quality)
See Appendix for detailed data.

In summary a number of points can be made regarding tubular system operation. The use of volume displacement rods is definitely not recommended. The price paid in headloss far exceeds the benefits obtained. If higher velocities are required, it appears recirculation of brine would be the preferred alternative. With regard to required velocities, it appears that a minimum velocity of about 1.5 fps is desirable, since operation at this velocity with the high flux-low salt passage modules was satisfactory. It should be noted, however, that the recoveries during this phase were quite low (45 - 70%), and that operation at higher recoveries may require higher velocities to offset the concentration polarization effects due to higher brine concentrations. Additional study at higher recoveries is necessary to answer this question. It is not known if the high initial flux losses experienced with the type 310 modules was specific for the modules utilized or a result of the lower salt rejection. In any event high salt rejection modules (greater than 99%) are definitely recommended for both flux and product water quality considerations.

Comparing the flux history in Figure 20 (310 modules) to the flux history from the previous field testing (3) significantly lower flux declines were noted in the present study. For example greater than 80% of the original flux was lost in 400 hours in the previous study compared to about 45% of the original flux in the present study. This would indicate that the pretreatment system did have some effect on the tubular system operation

Operation of the Hollow Fiber RO System

The operation of the hollow fiber RO system was accomplished concurrently with the tubular system using identical feed water. The initial permeator received at the site had an abnormally high salt passage (greater than 10%). This was a result of an improperly applied corrosion coating on the aluminum permeator shell, which resulted in poor brine flow distribution. This permeator was immediately replaced with a 316 stainless steel shell permeator and this corrected the salt passage problem.

The initial 838 hours of operation of the hollow fiber system were made using one permeator operating at 75% nominal product water recovery. Two additional permeators were then added to the system to form a 2-1 array (see Figure 14). This 2-1 array was operated for an additional 1832 hours. The flux history for the three hollow fiber modules is presented in Figures 22 and 23. Figure 22 is the flux history for permeator No. 6[illegible]. This permeator operated initially as a single unit at 75% recovery and then as the final stage in a 2-1 array. Figure 23 is the flux history for permeators 1129 and 1131 which were operated as the first stage of the 2-1 array. The permeators were operated in the 2-1 array to allow high recovery experiments and confine the expected $CaSO_4$ fouling to a single permeator. This also allowed

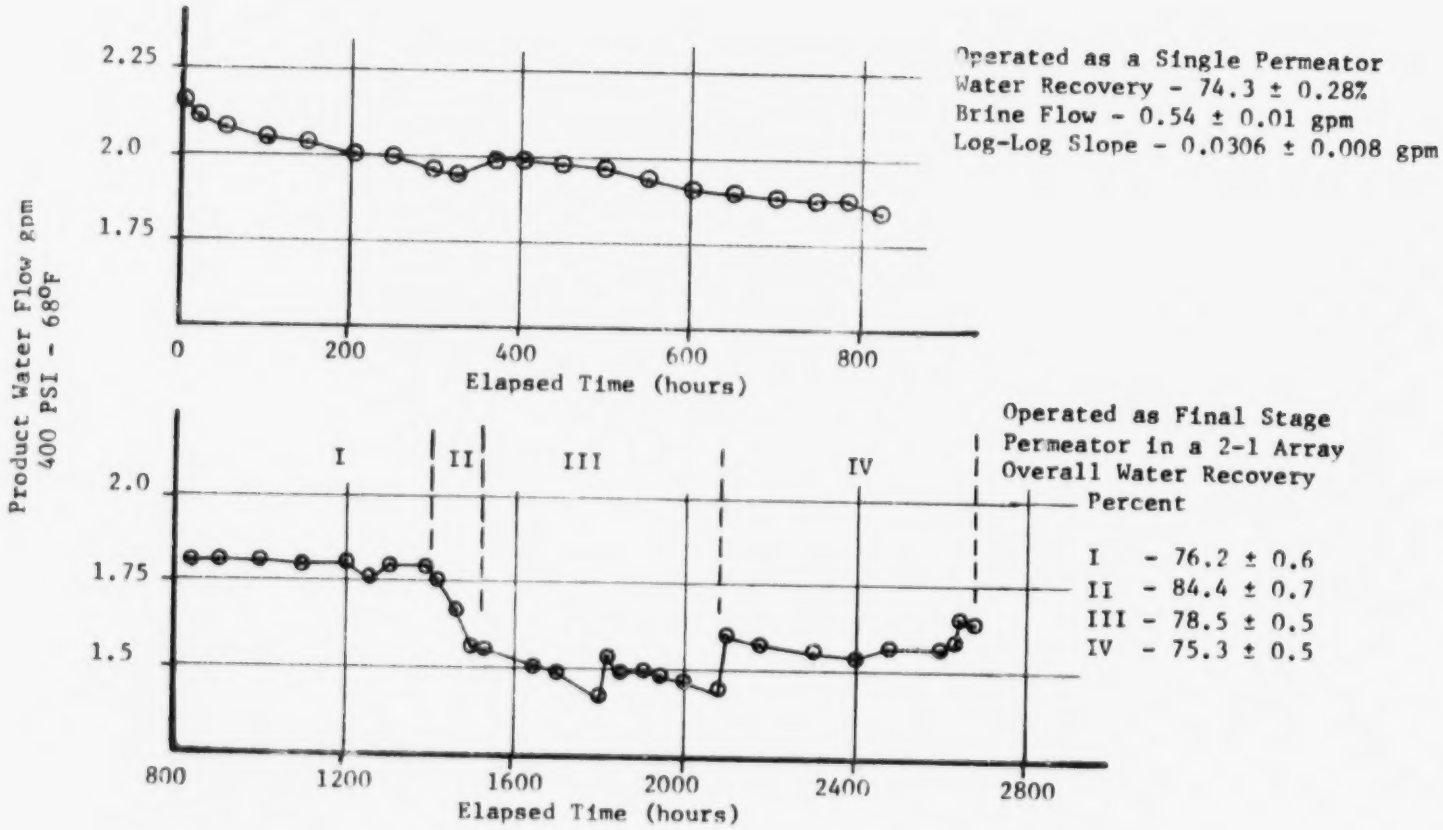


FIGURE 22

HOLLOW FIBER RO OPERATION PERMEATOR #691

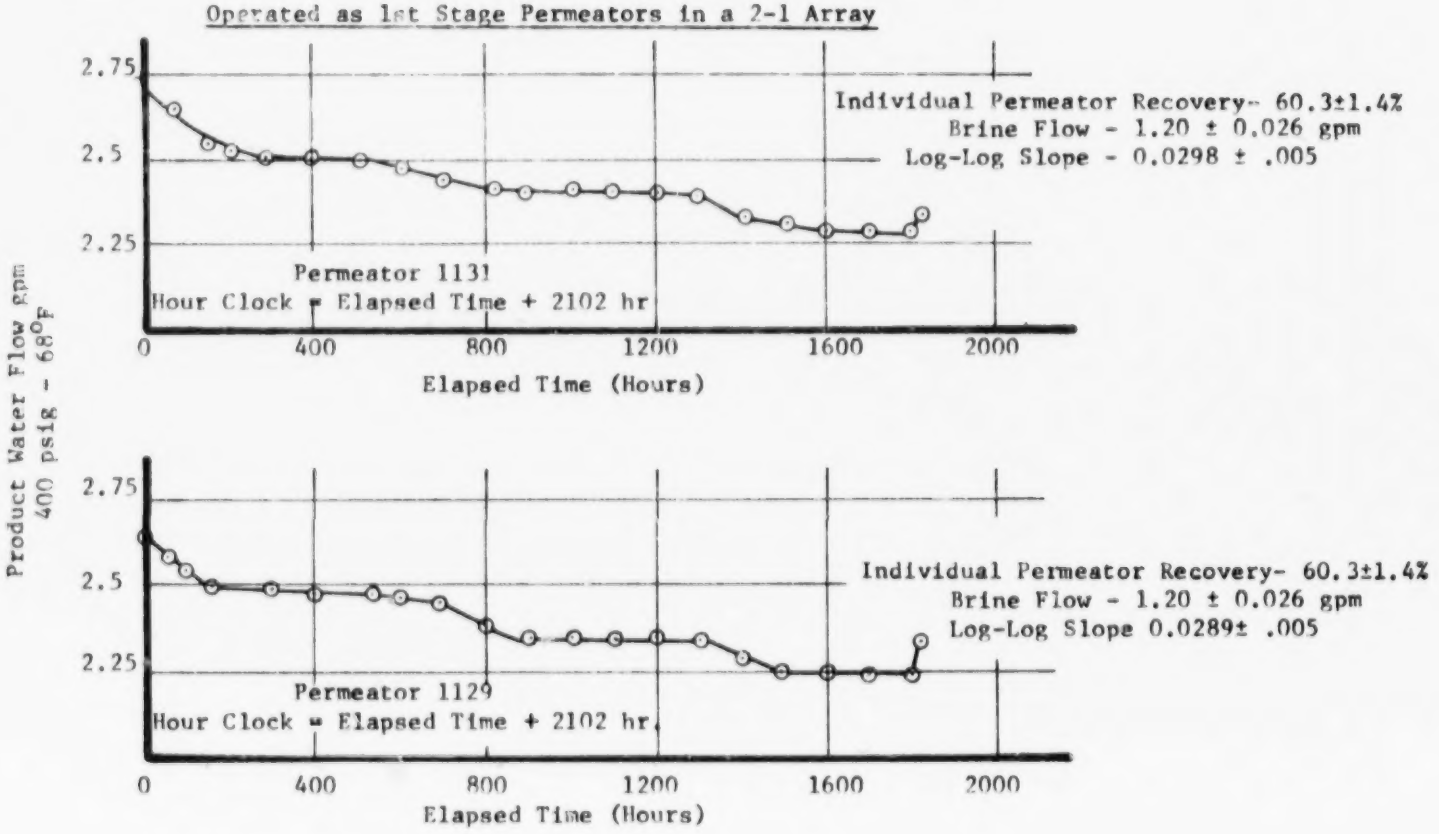


FIGURE 23

HOLLOW FIBER RO SYSTEM OPERATION

collection of long term flux data on the first stage permeators. Flow control orifices were utilized in the first stage permeators to insure equal flow of feed water to each permeator. A discussion of these orifices may be found in the Appendix on hollow fiber operational data. It should be noted that the hollow fiber system is not normally operated as a stage system, i.e., generally all permeators are operated in parallel. It should also be noted that the fluxes for the hollow fiber system were not listed as flow per unit membrane area, but as flow per permeator (module). This was done since the exact membrane area per permeator was not known.

The initial 838 hours of operation of the hollow fiber system (Figure 22) resulted in a relatively low flux decline. The log-log slope for this period of operation was 0.0306 ± 0.008 at the 95% confidence level. It should be noted that this slope was obtained by a regression analysis of 75 individual data points, all of which are not plotted in Figure 22. The correlation coefficient was 0.85 which indicates that all the flux decline is not due to membrane compaction. Membrane compaction is the loss of flux due to plastic flow (compaction) of the membrane. Pure compaction will plot a straight line on log-log paper and the correlation coefficient should be in the range of 0.95 to 0.97. Additional regression analysis on portions of the flux curve of Figure 22 indicate an increase in the log-log flux slope between 200 to 300 and 500 to 600 operating hours. This indicated fouling was occurring. An increase in the pressure drop across the permeator was also noted which would indicate fouling. This fouling was mainly due to iron precipitation, since calcium sulfate ($CaSO_4$) fouling does not generally cause an increase in pressure drop (see later discussion on $CaSO_4$ fouling), across the permeator, but rather an increase in salt passage. The water recovery during this period was 74.3 ± 0.28% and the brine flow 0.54 ± 0.01 gpm at the 95% confidence level.

After operating 838 hours with a single permeator, two additional permeators were added and operated in a 2-1 array. The flux history of these two new permeators is presented in Figure 23. The log-log flux slopes were 0.0298 ± 0.005 and 0.0289 ± 0.005 at the 95% confidence level. The regression analysis performed on this data also indicated fouling was occurring. Additional regression analysis on portions of the data indicated significant flux curve slope changes from 550 to 800 and 1300 to 1500 elapsed hours. Flux decline slopes outside of these time intervals closely approached the theoretical levels (high regression coefficients), indicating the fouling was occurring only for limited time periods and not continuously. The pressure drop across the bundle also increased during these periods of flux decline which would indicate iron fouling. Since the first stage permeators operated at about 60% water recovery calcium sulfate fouling was unlikely.

In order to evaluate the effect of the ultraviolet light on system operation, the light was turned off at 140 hours and remained off until 595 hours (Figure 23). This time interval represented 978 hours

to 1433 hours elapsed time on the last stage permeator (Figure 22). As may be seen in these figures, the fact that the UV light was off had little apparent effect on the flux rates. During this period, however, the pressure drop across the bundle did increase about 12 psi (from 18-30 psi). This fact would indicate some iron fouling was occurring. It is not known how much of the pressure drop was associated with the feed flow distributor. It is also interesting to note that during the time the UV light was off the second stage permeator had essentially no flux loss and no increase in bundle pressure drop. This would indicate that the first stage permeators were effectively filtering out the iron bacteria or other substance which was causing the increasing pressure drop across the first stage permeators.

When the two additional permeators were brought into operation, the original permeator was used as the last stage in the 2-1 array. Even though some iron fouling had occurred, the permeator (#691) was not flushed with any cleaning solutions. The flux history for the second stage operation of this permeator is presented in the lower curve of Figure 22. From 838 to 1400 hours the overall system recovery was 76.2 ± 0.6%. Essentially no flux was lost during this period (Phase I, Figure 22). At 1224 hours the overall recovery was increased to 85% and an immediate decline in flux was noted. The recovery was reduced to 75% at 1269 hours and the flux recovered to the same value as it was prior to the increase in recovery. The flux remained stable at this level until 1430 hours. If the high recovery period is ignored this period of time (838 to 1430 hours) represented a very stable period of operation showing essentially no flux decline for the last stage permeator. Brine flow rate during this period was 1.58 ± 0.03 gpm. During the same period of operation the flux slopes for the two permeators in the first bank were also stable with brine flows of 1.20 ± 0.026 gpm. A comparison of brine flows and flux decline slopes is shown below.

Case	Brine Flow (gpm)	Flux Decline Slope	Water Recovery (%)
1	0.54 ± 0.01	0.0306 ± 0.008	74.3
2	1.20 ± 0.026	0.012 ± 0.006	60.3
3	1.58 ± 0.03	0.011 ± 0.004	76.2

Based on the above comparison there appears to be a correlation between flux decline and brine flow. It should be noted, however, that Case 1 was over a different time period than Cases 2 and 3, and an absolute comparison cannot be made. In view of the large differences in decline rates it is recommended that a minimum brine flow of about 1 gpm be maintained.

At 1450 hours the overall system recovery was again raised to 85%. The flux immediately began to decline at a rapid rate (Figure 22) in the second stage permeator. The salt passage also increased, but the

head loss across the bundle remained constant. This was a definite indication of $CaSO_4$ fouling. The $CaSO_4$ precipitation occurs in the outer most fibers, since the brine is most concentrated in this area. A disruption in brine flow distribution results in some areas of the fibers receiving no brine flow. This causes an increased salt passage. The brine now short circuiting around these areas finds the path of least resistance and this results in little change in head loss across the fiber bundle. However, if the condition persists for a long enough time an increase in headloss would be noted as a greater and greater portion of the brine flow area is plugged. During this same period (595 - 668 hrs., Figure 23) the flux decline slope increased for the two first stage permeators. The brine flows were also reduced to about 1.0 gpm during the period as compared to 1.2 gpm prior to increasing the system recovery. This decrease may have initiated the flux decline noted in the first stage permeators.

At 1537 hours (Figure 22), a fifteen minute high brine flow flush was performed on the second stage permeator and the water recovery was lowered to 80%. The flux decline rate decreased, but was still significantly higher than experienced at 75% water recovery. At 1700 hours the last stage permeator was flushed in an attempt to remove some of the $CaSO_4$ precipitation which had occurred. A 50 gallon (2 wt %) solution of ammoniated citric acid was recirculated through the permeator for 2 hours at a pH of 4.0. A new solution was then made and buffered to pH 8.3 and this solution was recirculated for 2 hours. This flushing was only marginally successful and recovered only about 25% of the flux lost due to $CaSO_4$ fouling. The system was put back into operation and operated at 80% recovery for an additional 250 hours. The flux decline slope was approximately the same as before the flush, and still significantly higher than the 75% recovery level.

At 2100 hours the last permeator was flushed with a 3.4 wt % EDTA and 1.7 wt % $Na_2S_2O_4$ solution for a period of two hours. This flush recovered about 50% of the total flux lost due to $CaSO_4$ fouling. Next a 2 wt % sulfamic acid flush was used, but this did not recover any additional flux. At this point it was decided to wait until the end of the study before attempting any additional cleaning on the second stage permeator. The system was put back on line and operated another 600 hours at an overall recovery of 75.3 ± 0.5%. The flux over this period was very stable as shown in Figure 22 from 2100 to 2700 hours.

The flux history at higher recoveries provided some valuable information with regard to $CaSO_4$ fouling. Figure 24 presents a plot of the $CaSO_4$ solubility product experienced in the brine at various time intervals and compares these values with the flux history over the same period.

The peaks of the $CaSO_4$ curve correlate well with the high flux decline periods. The curves in Figure 24 indicate a molar solubility product of about 25 to 35 x 10^{-5} is all which can be attained without $CaSO_4$ fouling. Above this range $CaSO_4$ fouling will cause rapid flux declines.

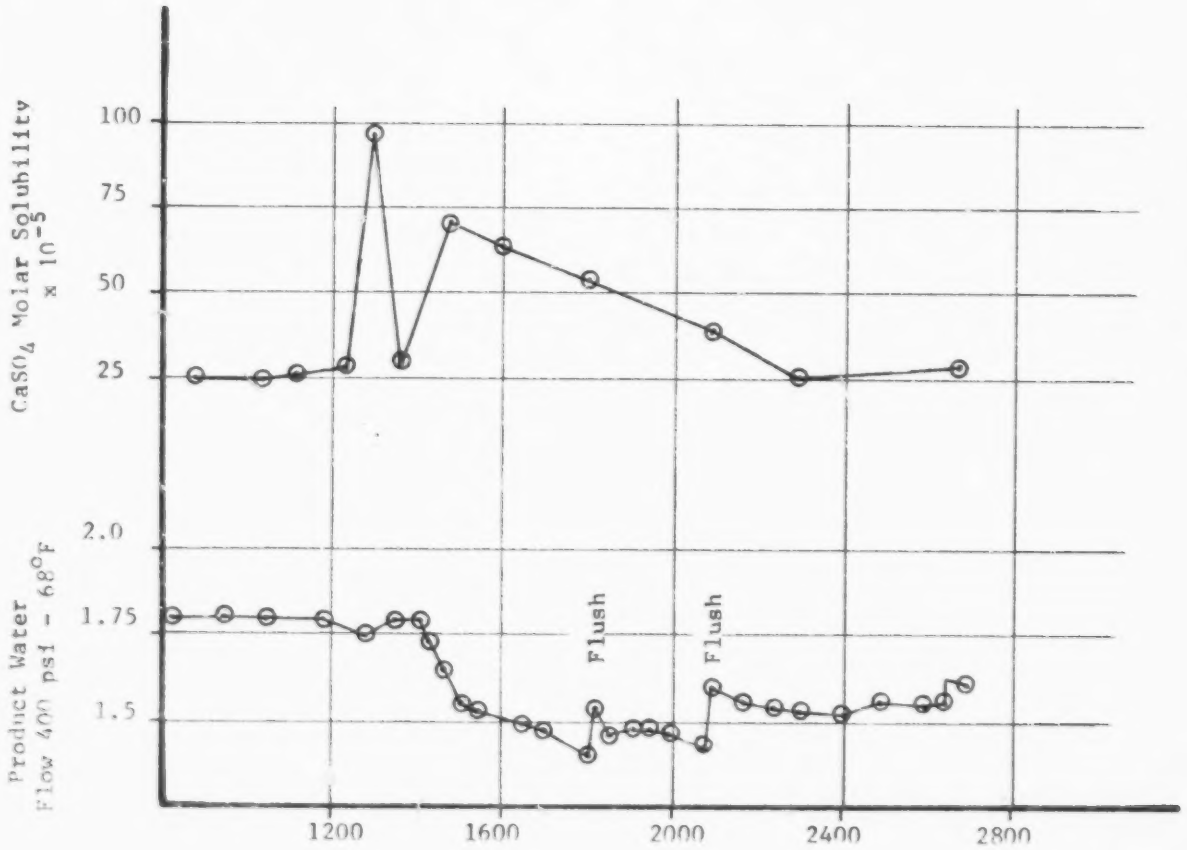


FIGURE 24

COMPARISON OF $CaSO_4$ MOLAR SOLUBILITY PRODUCT AND PRODUCT WATER FLOW

The conclusion can then be made that $CaSO_4$ levels in the raw feed water will determine the maximum level of product water recovery.

Operation of the hollow fiber unit was terminated at total operating hours of 2670. At this time the entire unit was flushed with a 4 wt % solution of $Na_2S_2O_4$. This caused a flux increase in all permeators as shown in Figures 22 and 23, but did not restore the bundle pressure drop to the values experienced at the start of the study. This indicates that the flushing was not completely successful in removing the iron fouling. The units were then filled with product water and disinfected with formaldehyde.

A summary of the chemical analysis for the hollow fiber unit is presented in Table 13. Average brine and product water quality and the range at the 95% confidence levels are shown. Salt rejection percentage is presented for all three permeators. In general, salt rejection was in the range of 99.2 to 99.7% with the exception of silica which was 94-98%. Total dissolved solids rejection was about 99%. It can be noted that the product water produced even at these high salt rejections does not meet USPHS standards with regard to iron and manganese which are 0.3 mg/l and 0.05 mg/l respectively. The pH is also too low at 4.2. The product water would therefore require additional treatment. If the product water were blended with an existing drinking water supply it would probably meet the requirements. If it is desired to utilize the product water directly the iron and manganese levels will have to be reduced and sufficient buffering capacity added to provide a stabilized water. This buffering capacity is extremely important since the water is almost equal to distilled water and could cause corrosion problems.

A single set of analyses was performed to determine the carbon dioxide levels in the feed, product and brine flows of the hollow fiber system. These analysis indicated the raw feed water had a CO_2 level of 6.4 mg/l. The product water contained 6.9 mg/l while the brine only contained 2.9 mg/l. Mass balances across the system were correct within 1.5%. This means that essentially all the CO_2 was passing through the membrane i.e. no rejection of CO_2 was being observed. The test prodedure consisted of purging the CO_2 from the mine water by air stripping. The CO_2 laden air was bubbled through a potassium chromate-sulfuric acid solution, through an iodine solution and finally through barium hydroxide. The CO_2 caused a precipitate of barium carbonate to form and the CO_2 concentration was determined by back titration of the barium hydroxide solution.

In summary the operation of the hollow fiber system was satisfactory since only a slight amount of iron fouling was experienced. This fouling can be controlled by periodic flushing of the membranes or by addition of acid to lower the feed pH as suggested by others (8). Acid addition will eliminate iron fouling by keeping all ferric iron in the soluble state. Calcium sulfate fouling can be controlled by keeping the recovery at the proper levels. It may be concluded that

TABLE 13

SALT REJECTION CHARACTERISTICS
HOLLOW FIBER RO SYSTEM

Ion	Raw Water Quality mg/l	Brine Quality mg/l	Product Water Quality mg/l	Salt Rejection % Module 1129	Module 1131	Module 691
LABORATORY ANALYSIS						
Calcium	111 ± 6	487 ± 69	0.76 ± 0.2	99.67 ± 0.05	99.66 ± 0.07	99.69 ± 0.09
Magnesium	83 ± 4	381 ± 56	0.66 ± 0.1	99.67 ± 0.06	99.64 ± 0.08	99.65 ± 0.07
Manganese	14 ± 0.6	57 ± 10	0.12 ± 0.03	99.66 ± 0.16	99.61 ± 0.08	99.61 ± 0.12
Iron (Total)	70 ± 6	308 ± 60	0.60 ± 0.2	99.60 ± 0.16	99.43 ± 0.26	99.62 ± 0.11
Aluminum	8 ± 0.3	35 ± 5	0.21 ± 0.09	99.34 ± 0.05	99.67 ± 0.08	99.30 ± 0.3
Silica	11 ± 0.5	49 ± 6	0.83 ± 0.3	94.40 ± 1.9	94.20 ± 1.9	97.66 ± 0.6
Total Dissolved Solids	1319 ± 85	5809 ± 862	25 ± 9	99.10 ± 0.4	98.95 ± 0.7	99.35 ± 0.3
FIELD ANALYSIS						
Calcium	118 ± 6	523 ± 53	1.04 ± 0.2	99.74 ± 0.11	99.76 ± 0.11	99.63 ± 0.08
Total Hardness As ($CaCO_3$)	602 ± 42	2656 ± 274	5.0 ± 1	99.77 ± 0.06	99.77 ± 0.71	99.59 ± 0.10
Iron (Total)	67 ± 5	309 ± 40	0.60 ± 0.06	99.78 ± 0.03	99.76 ± 0.04	99.63 ± 0.04
Iron (II)	64 ± 4	280 ± 37	0.57 ± 0.06	99.77 ± 0.03	99.75 ± 0.04	99.62 ± 0.05
Sulfate	774 ± 45	3146 ± 381	3.2 ± 1.0	99.90 ± 0.06	99.90 ± 0.04	99.80 ± 0.06
pH	3.38 ± 0.06	2.8 ± 0.02	4.2 ± 0.1	--	--	--

NOTES: All ranges at 95% confidence level
Salt Passage (%) = 200 (product water quality) /(feed water quality + brine quality)
See Appendix for detailed data

the feasibility of acid mine treatment by RO has been established.

The mechanical operation of the hollow fiber system was excellent. No failures were experienced. With regard to the remainder of the RO components, the only item which required maintenance was the RO high pressure pump. Frequent greasing was required for the pump packing. No shear pan failures were experienced in the entire 3800 hour run since the pins were replaced four times during the run. The system was never shut down longer than 15 minutes for pump maintenance.

Operation of the Brine Treatment System

The brine exiting from the RO units contains all the impurities originally present in the raw waste. The main constituents include calcium, magnesium, manganese, iron, aluminum, silica, and sulfate. The concentration of these elements will be related to the feedwater quality and the water recovery of the RO unit (see Table 13). In any brine treatment system, the objective is removal of these pollutants. In the case of iron and aluminum this is easily accomplished, since these metals are quite insoluble in certain pH ranges and can be precipitated as the metal hydroxides. Magnesium on the other hand is quite soluble and difficult to remove. Manganese when oxidixed to the travelent state will also form insoluble hydroxides, but complete removal can be obtained only at a pH above about 9.5. Silica will also complex with calcium at this high pH and be removed from the liquid phase. Calcium sulfate will precipitate to a limited extent in accordance with its solubility, however, the overflow from a neutralization system will be saturated with respect to $CaSO_4$. Since many of the impurities can be removed by simple neutralization, the brine treatment system consisted of neutralization followed by aeration to oxidize the iron and possibly manganese, and then sedimentation. The brine treatment system was operated on either the tubular RO brine or the hollow fiber brine, but not a mixture. Bench scale tests were run in conjunction with the full scale system to adequately define the operating criteria which were utilized.

Bench Scale Testing

Iron oxidation tests were performed at pH 6.6 and 7.7 to determine the iron oxidation rates. Figure 25 presents the data collected. It is obvious that a pH of about 7.7 is required for rapid and complete oxidation of the ferrous iron. The source of water for this test was RO brine from the hollow fiber unit operating at 75% recovery.

Bench scale settling rate tests were performed at various oxidation times. The results of these tests are plotted in Figure 26. It may be seen that the settling rate increases with aeration time at a constant pH of 7.7. It is not known if higher settling rates would be produced at lower oxidation times by raising the pH. This is an area which required additional study. In any event the settling rates at

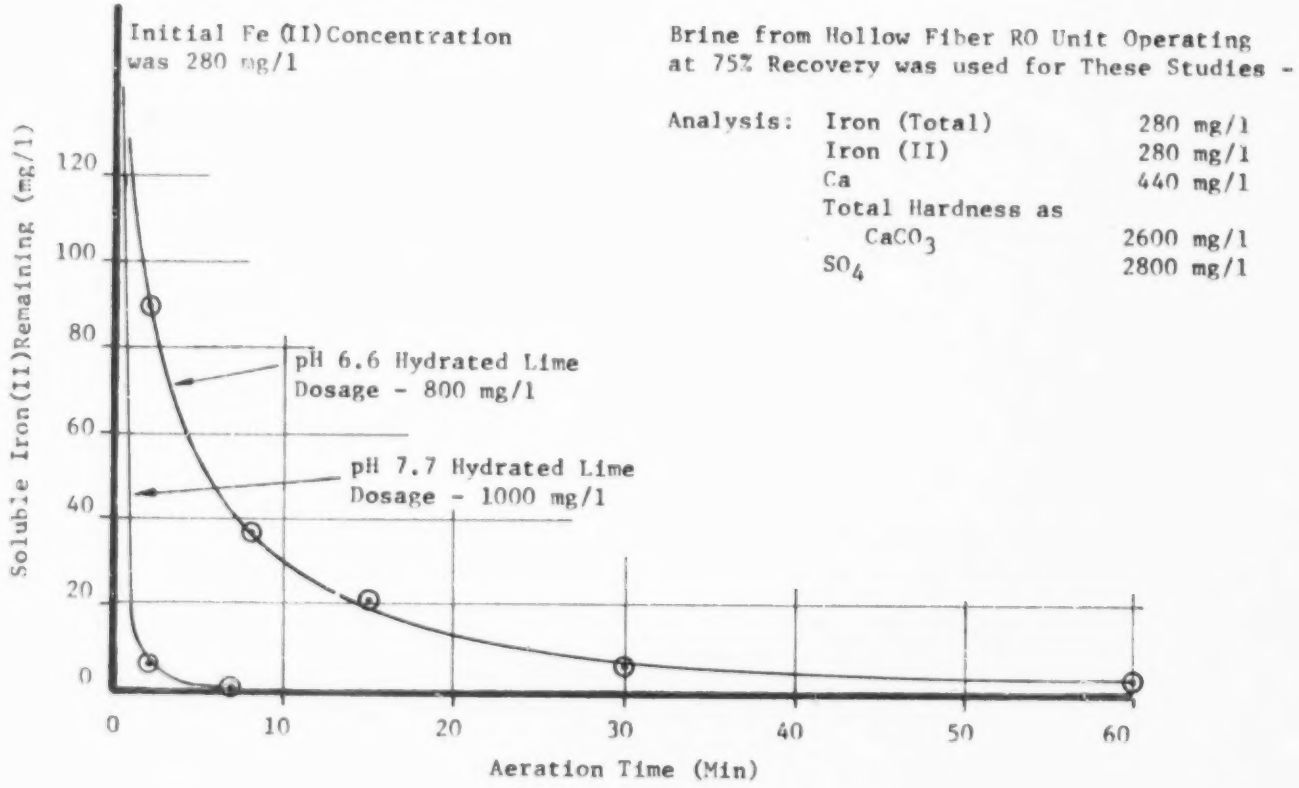


FIGURE 25

IRON OXIDATION STUDY

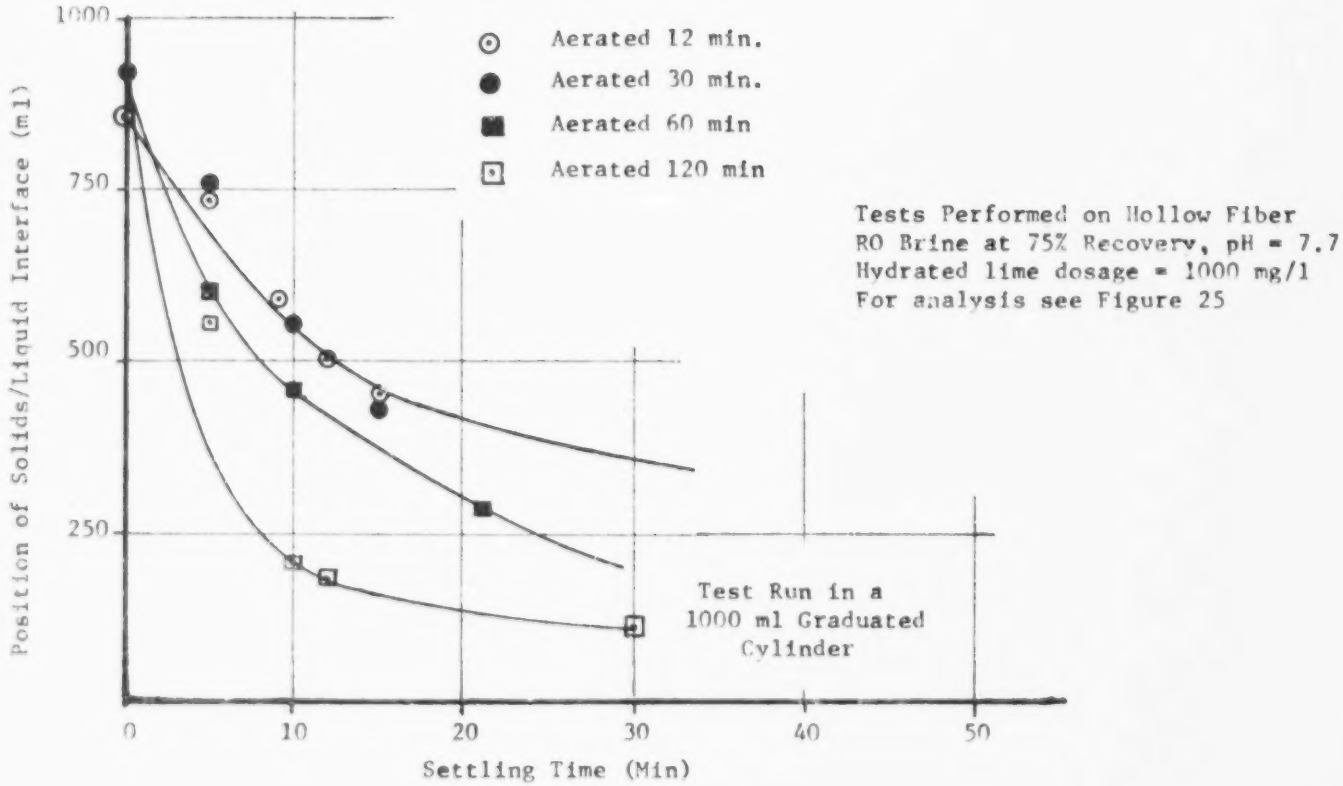


FIGURE 26

SETTLING RATE TESTS

60 and 120 minutes of aeration provide sufficient settling velocity for reasonable clarifier overflow rates.

A test was also run where the iron was not oxidized. The pH was raised to 7.7 and the waste flocculated and settled. This procedure produced a settling rate comparable to the curve for 60 minutes aeration (Figure 26), however, the effluent contained 70 mg/l of soluble iron (II). It was, therefore, concluded that this mode of treatment was not feasible.

In order to evaluate neutralization on a bench scale, two experiments were run to evaluate $CaSO_4$ precipitation, which might occur in the system. The results of these two tests are shown in Table 14. Both tests were run in a 1000 ml graduated cylinder. Test I utilized 75% RO brine. The lime ($Ca(OH)_2$) dosage was 1288 mg/l which was equivalent to 695 mg/l of calcium. As may be seen in Table I, the effluent contained 800 mg/l Ca after 22 hours of sedimentation. This was 360 mg/l higher than the influent brine. Considering the calcium added when neutralizing the brine, a net of 335 mg/l of calcium was precipitated. The sulfate analysis, however, remained constant. The calcium could have been precipitated as a calcium silicate complex and as calcium carbonate (from the CO_2). It is also possible that all the lime did not dissolve. It was obvious however, that neutralization and settling produced a water higher in calcium when compared to the RO brine.

Bench scale Test II (Table 14) was run on 85% recovery RO brine. Lime ($Ca(OH)_2$) dosage was 1800 mg/l or 974 mg/l calcium. This test lasted 38 hours. During this time a steady decrease in calcium was found in the effluent. Sulfate also decreased indicating $CaSO_4$ was precipitating. Total calcium in the feed considering the lime added was 1656 mg/l which means 816 mg/l precipitated. The sulfate precipitated was 2200 mg/l. This would account for an equivalent of 915 mg/l calcium. The possible error in the sulfate analysis (±10%) could account for this difference. In any event, the calcium in the effluent was still higher than the influent brine.

Full Scale System Operation

The full scale neutralization system was operated on the RO brines for four runs. Tables 15 and 16 summarize the operation of the neutralization system. The results confirmed the bench scale tests. Good removals of iron, and aluminum were achieved (95 to 99%). Iron was removed as iron hydroxide and aluminum as aluminum hydroxide. Silica removals were also high (95%) and removal was most probably via a calcium silicate complex. Manganese removals were lower in the range of 55 to 80% as manganese hydroxide. Calcium on the other hand increased as previously noted in the bench scale studies (due to lime addition for neutralization). Manganese was not removed to any extent and in fact exhibited an increase in one case. It may be

TABLE 14

BENCH SCALE NEUTRALIZATION TESTS

Test I	Feed	Effluent	
		30 Min	22 Hrs
Total Iron, mg/l	296	5.8	0.8
Iron (II), mg/l	280	0	0
Sulfate, mg/l	2950	3000	3000
Total Hardness ($CaCO_3$) mg/l	2600	3200	3200
Calcium, mg/l	440	800	800

Test Specifications

Lime Dosage - 1288 mg/l $Ca(OH)_2$ = 695 mg/l Ca

pH - 7.9

Brine from Hollow Fiber RO Unit - 75% Recovery

Temperature 73° F

Test in 1000 ml graduate

Test II	Feed	Effluent			
		30 Min	12 Hrs	21 Hrs	38 Hrs
Total Iron, mg/l	505	4.3	0	-	-
Iron (II)	450	0	0	-	-
Sulfate, mg/l	6250	5000	5000	4750	4050
Total Hardness $CaCO_3$, mg/l	3900	5100	4900	4650	4150
Calcium, mg/l	682	1180	1180	1040	840

Test Specifications

Lime Dosage - 1800 mg/l $Ca(OH)_2$ = 974 mg/l Ca

pH - 7.9

Brine from Hollow Fiber RO Unit - 85% Recovery

Temperature 74° F

Test in 1000 ml graduate

TABLE 15

SUMMARY OF NEUTRALIZATION SYSTEM OPERATION

Run #	1	2	3	4
Brine Flow Rate, gpm	2.5	1.7	0.88	1.7
Duration of Run, Hrs.	73	118	37	46
Aeration Time, Hrs.	1.95	2.86	7.2	2.86
Settling Time, Hrs.	11.5	17.0	32.8	11.5
Hydrated Lime Dosage, mg/l $(Ca(OH)_2$	505	1215	1720	1215
INFLUENT WATER QUALITY				
pH	2.9	2.8	2.8	2.9
Iron, mg/l	108	287	475	340
Calcium, mg/l	368	440	720	560
Total Hardness, mg/l as $CaCO_3$	1580	2950	4000	2600
Sulfate, mg/l	1400	2500	6500	3100
EFFLUENT WATER QUALITY				
pH	6.5-7.0	7.8	7.6	7.7
Iron	6.0	4.5	0.9	5.5
Calcium, mg/l	340	800	1080	800
Total Hardness as $CaCO_3$, mg/l	1430	3350	5000	3700
Sulfate, mg/l	1250	2650	4750	3500

concluded that the quality of the effluent from the neutralization system operating at a pH of 7.6-7.8 eliminated the possibility of recycling this water to the RO system (in the light of the high calcium content) as suggested by Hill et al. (9). It is possible that operation at a lower pH would allow recycling, but clarity of the settling tank overflow fould be adversely effected causing additional treatment problems. It is also possible that other treatment systems could be utilized to render the sedimentation tank overflow amenable to recycling back through the RO unit and hence eliminate this liquid waste stream.

TABLE 16

SUMMARY OF ATOMIC ABSORPTION ANALYSIS OF NEUTRALIZATION SYSTEM

Neutralization Run #	3		4	
	Influent	Effluent	Influent	Effluent
Calcium, mg/l	715	1030	500	800
Magnesium, mg/l	572	990	376	361
Manganese, mg/l	86.4	38.8	55.9	29.9
Iron, mg/l	493	0.3	312	1.4
Aluminum, mg/l	54	0.1	25	0.1
Silica, mg/l	76	4	45	3

SECTION VIII

GENERAL DISCUSSION

Discussion of Flushing Techniques

Two sources of fouling were experienced during the field operation, i.e. iron and $CaSO_4$. Flushing methods were evaluated for effectiveness in removing the precipitates from the RO modules and the effects of the flushing solution on the RO membranes.

For removal of iron fouling, a two-weight percent product water solution of citric acid adjusted to a pH of 4 with ammonia was evaluated. This solution did not effectively dissolve the iron from the membrane, as little flux changes were experienced. To further evaluated this solution some precipitated iron was scraped from a raw AMD storage tank and put into the ammoniated citric acid. The sample was then mixed on a magnetic stirrer for 2 hours. Visual inspection revealed little if any iron precipitate had dissolved. This same experiment was performed using sodium hydrosulfite ($Na_2S_2O_4$) in a 4 weight percent solution and the results were dramatic. All traces of iron precipitate were dissolved within 15 minutes. This is consistent with results from a previous study (3). The next step was to use sodium hydrosulfite on an actual RO module. The 610 tubular modules which had been fouled with iron were flushed with a 4 weight percent solution for 1.5 hours. This resulted in an increase in flux from 12 gsfd to 17 gsfd. After operating the system for about 2 hours the salt passage was at the same level as prior to the flush. The hollow fiber modules were then flushed with sodium hydrosulfite; flux increases were also experienced from 2.26 to 2.42 gpm per module for the first stage modules. Salt passage returned to normal after 20 hours of operation. Based on the above testing it appears that sodium hydrosulfite is an effective method for cleaning AMD-iron fouled RO membranes.

For removal of $CaSO_4$ fouling the following solutions were evaluated: a 2 wt % solution of citric acid buffered to pH 8 with ammonia; a 3.4 wt % solution of EDTA - 1.7 wt % solution of $Na_2S_2O_4$; and a 2 wt % solution of sulfamic acid. The only module fouled with $CaSO_4$ was the hollow fiber module No. 0691. (This was the only module operated at high recovery.)

Flushing in the field was first accomplished utilizing the ammoniated citric solution at pH 8 for 2 hours. This resulted in restoration of about 25% of the flux which had been lost due to $CaSO_4$ fouling. Analysis of the flushing solution indicated an increase of calcium levels from 19 mg/l to 870 mg/l indicating calcium was being brought into solution. The module was next flushed using the EDTA - $Na_2S_2O_4$ solution. This resulted in restoration of flux from 1.4 to 1.6 gpm, while the flux prior to $CaSO_4$ fouling was 1.8 gpm. The final flushing

in the field was accomplished using the sulfamic acid solution for a period of two hours. This resulted in no appreciable increase in flux. No additional field flushing was attempted. At the completion of the study, permeator 691 was returned to the laboratory for additional flushing. Based on flush water analysis the best solution appeared to be ammoniated citric acid at pH 8. It was felt that the time of flushing was important and that the limited success in the field with this solution was a result of insufficient flushing time. Prior to flushing, the module was recharacterized and then the bundle was removed for inspection. A solid ring of $CaSO_4$ precipitate was found near the brine exit end of the module. The bundle was reinstalled and the flushing precedure along with performance results are presented in Table 17. As may be seen the precipitate was difficult to remove, however, after 20 hours of flushing the performance was restored to the same levels that existed in the field prior to the high recovery run and subsequent $CaSO_4$ fouling. Inspection of the bundle indicated no traces of the calcium sulfate deposits previously noted. The fiber bundle was then unrolled and all fibers inspected. No pockets of precipitates were found in the entire bundle. Tests were then run on the fiber to determine if any damage had occurred during 2670 hours of operation and the many chemical flushes which were performed. Fiber strength, and elasticity indicated absolutely no damage had occurred to the fibers as all tests were comparable to new fiber (14). It was concluded that $CaSO_4$ fouling could be removed utilizing ammoniated citric acid at pH 8 and that no membrane damage was observed from utilization of the various flushing solutions or operation on AMD for 2670 hours.

Economic Consideration for RO - AMD Operation

Based on the results of the field evaluation phase, estimates of the costs associated with treatment of AMD via RO were prepared. The flow sheet utilized is shown in Figure 27. The following assumptions were made to arrive at the costs shown.

1. Hollow fiber RO modules were utilized.
2. RO product water capacity was 750,000 gpd.
3. Chemical additive costs were based on field testing results.
4. Diatomaceous earth filtration was utilized.
5. No costs for buildings or land were included.
6. The product water from the plant meets USPHS standards.
7. No costs were included for disposal of residuals.
8. Operating manpower included a plant manager and a crew of 3. Total salary and administrative costs - $50,000 per year.

TABLE 17

LABORATORY RO MODULE CLEANING RESULTS

Test Sequence - Description	Product Water Flow 400 psi-68°F gpm	Water Recovery %	Salt Passage %	Bundle ΔP
Original performance data prior to field operation	2.17	76.4	2.7	0
Performance upon returning from field	1.64	74.8	11.6	9
Performance after 3 hours shell feed flush	1.82	74.7	10.7	11
Performance after 3 hours shell feed - 3 hours distributor feed flushes	1.98	75	9.4	8
Performances after 5 hours distributor feed flush - 14 hour tap water flush 6 hour shell feed flush	1.85	75.6	3.5	3.5

NOTES: Flushing solution ammoniated citric acid pH 8.
Test solution 1500 mg/l NaCl

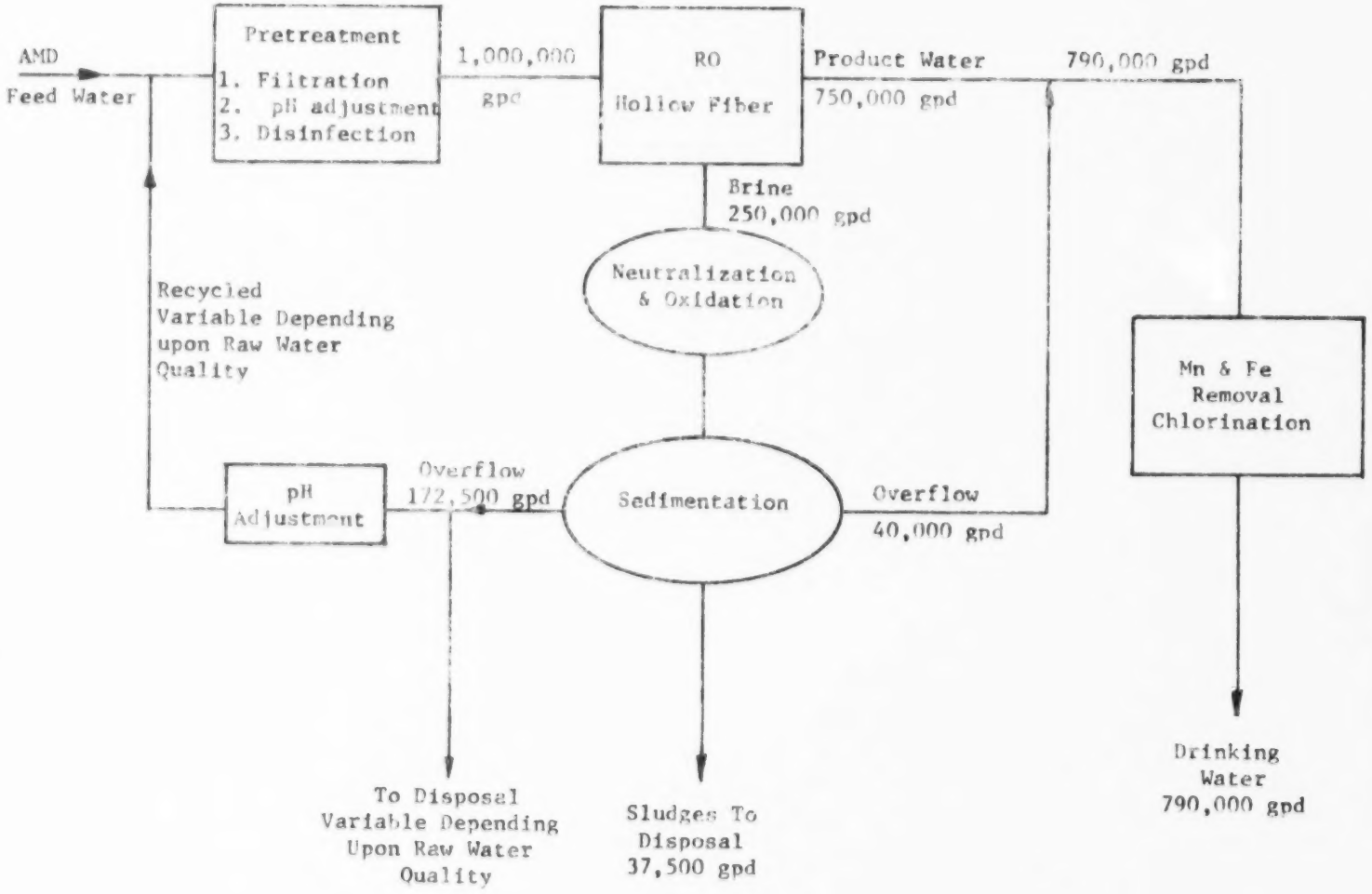


FIGURE 27

FLOW SHEET USED FOR COST ESTIMATES

9. Power costs at 1.0 ¢/kwh

10. Chemical additives include acid, diatomaceous earth, lime, chlorine, flushing chemicals for RO membranes, potassium permanganate.

11. RO module life 4 years - replacement cost 28¢/gpd capacity.

12. Brine treatment system of concrete construction with high speed floating aerators.

13. Product water treatment system utilizes a portion of the sedimentation tank overflow for neutralization and potassium permanganate for manganese oxidation - followed by filtration and chlorination.

Shown below are the major cost items for the treatment system of Figure 27. All cost estimates are based on vendor quotations or recent purchase prices.

I. Capital Costs

A. Pretreatment
   Filtration (diatomaceous earth)
   pH control
   Disinfection . . . . . $ 29,000

B. RO System
   Modules
   Pumps and plumbing
   Instrumentation. $385,000

C. Brine Treatment System
   Aeration unit (high speed surface aerator)
   Sedimentation unit
   Chemical feeders and controls. . . . . $ 58,000

D. RO Product Water Treatment
   Iron and Mn removal
   Final filtration
   Chlorination . . . . . $ 31,000

TOTAL CAPITAL COST $503,000

Amortized @ 6% - 20 yr = 15¢/1000 gal Product Water

II. Operating Costs in ¢/1000 gal Product Water

A.	Chemical additives	4.8
B.	RO Modules	17.4
C.	Power	7.0
D.	Maintenance - Materials	2.0
E.	Operating Manpower	17.3
	TOTAL	48.5

The costs presented herein are estimated based on present day prices. Advancement in RO hardware will undoubtedly bring price reductions in the RO equipment. Also refinement of the flow sheet may also result in more economical operation. One must also consider that two tasks are being performed i.e. waste treatment and production of potable water.

SECTION IX

ACKNOWLEDGEMENTS

Many people in the Rex Ecology Division contributed to the success of this project. Design of the demonstration system was made by F. Toman with assistance from J. E. Milanowski. Operation of the unit was performed by F. Toman, M. K. Gupta, and D. G. Mason. Laboratory analyses were provided by the Ecology Division Analytical Laboratory headed by R. E. Wullschleger. Bench scale experiments were conducted by M. K. Gupta. The report was written by D. G. Mason and M. K. Gupta.

Assistance from the staff at the EPA Acid Mine test site, Norton, West Virginia is appreciated. Review of RO operations with Robert Scott and Roger Wilmoth provided valuable information. Alvin Irons and Randolph Lipscomb provided expert assistance in field operation of the RO unit.

Guidance from the EPA project officer, Ronald Hill, and key State of Pennsylvania personnel, David Maneval and John Buscavage, is also appreciated.

SECTION X

REFERENCES

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SECTION XI

PUBLICATIONS

Portions of the work described herein will be utilized in a technical paper to be presented at the 4th Symposium on Acid Mine Drainage Mellon Insitute, Pittsburgh, Pennsylvania in April 1972.

SECTION XII

GLOSSARY OF TERMS

AMD - Acid Mine Drainage

Brine - The waste stream exiting a Reverse Osmosis unit, also can be called concentrate

Concentrate - The waste stream exiting a Reverse Osmosis unit, can also be called brine

Flux - Rate of water passage through a Reverse Osmosis membrane usually expressed as gallons per sq ft of membrane per day (gsfd)

Permeate - Water which has passed through a Reverse Osmosis membrane - could be termed product water

Permeator - A hollow fiber Reverse Osmosis module, trade mark of DuPont

Product Water - Water which has passed through a Reverse Osmosis membrane - could be termed permeate

RO - Reverse Osmosis

Salt Rejection - Measure of the amount of salts not passing through the membrane

Salt Passage - Measure of the amount of salts passing through the membrane.

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SECTION XIV
APPENDICIES

I. Operating Data Hollow Fiber RO System

This Appendix contains all operating data for the hollow fiber RO system. The normalized flux values were calculated using the temperature and pressure correction equations shown below:

For Temperature - $Q_{68} = Q_T \times (1.01667^{68-T})$

where: Q_{68} = Product water flow at $68^{o}F$

Q_T = Product water flow at observed temperature ^{o}F

T = Temperature ^{o}F

For Pressure - $Q_{400} = Q_o\ [800/(P1 + P2)]$

where: Q_{400} = Product water flow at 400 psig

Q_o = Observed product water flow

P_1 = Pressure into the module, psi

P_2 = Pressure out of the module, psi

When the system was operated as a 2-1 array flow control orifices were utilized to insure equal flow distribution to the two first stage permeators. The orifice headloss equation is:

$\Delta P = Q_B \times 22.2293$

where ΔP = the head loss across the orifice in psi

Q_B = the flow through the orifice

The orifices were located in the brine line ahead of the pressure gauge, hence, the interstage pressure shown in Table I-2 includes the orifice pressure drop.

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TABLE I-1

HOLLOW FIBER OPERATING DATA
FOR SINGLE PERMEATOR

Elapsed Time (hrs)	Pressure In (psi)	Pressure Out (psi)	Temp. °F	Brine Flow gpm	Product Flow sec/gal	Normalized[1] Product Flow gpm	Product Water Recovery - %
0	400	400	53.0	0.50	35.50	2.166	77.2
3	400	400	53.2	0.51	35.60	2.153	76.9
9	400	400	52.3	0.51	36.20	2.149	76.5
19	400	400	52.3	0.51	36.40	2.137	76.4
21	400	400	53.2	0.51	36.10	2.123	76.5
25	400	390	54.0	0.55	36.60	2.092	74.9
25	400	390	54.0	0.54	36.50	2.098	75.2
32	400	390	53.8	0.43	37.00	2.077	74.9
43	400	390	53.0	0.54	37.40	2.082	75.0
49	400	390	53.0	0.53	37.30	2.087	75.1
52	400	390	52.0	0.54	37.30	2.087	75.0
68	400	389	52.9	0.53	37.60	2.077	74.9
73	400	390	54.0	0.53	37.40	2.048	75.0
81	400	390	53.0	0.54	37.60	2.071	74.8
94	400	390	53.0	0.54	38.10	2.044	74.4
95	400	400	53.0	0.49	37.20	2.067	73.2
101	400	400	53.0	0.59	37.40	2.056	73.1
114	400	400	52.0	0.58	39.40	1.955	72.5
118	380	380	53.0	0.58	38.90	2.081	72.5
119	400	400	53.0	0.60	37.20	2.061	73.0
125	398	398	53.0	0.59	38.00	2.034	72.7
137	400	400	52.9	0.59	36.90	2.087	73.3
146	400	400	53.0	0.58	36.70	2.095	73.7
151	380	380	53.0	0.58	37.80	2.141	73.1
163	400	400	53.0	0.58	38.10	2.018	73.0
188	400	400	53.0	0.59	38.00	2.023	72.9
198	400	400	53.0	0.59	38.10	2.018	72.9
211	400	400	53.0	0.58	38.50	1.997	72.8
216	400	400	53.0	0.58	38.75	1.984	72.7
222	400	400	53.0	0.58	38.90	1.977	72.6
235	400	400	53.0	0.58	38.60	1.992	72.9
240	400	400	54.0	0.58	38.60	1.959	72.9
243	400	400	54.0	0.59	38.35	1.972	72.7
245	400	400	53.0	0.58	38.85	1.979	72.5
247	400	400	53.0	0.59	38.10	2.018	72.7
257	400	400	53.0	0.60	38.50	1.997	72.2
263	400	400	53.5	0.52	38.45	1.983	74.9
284	400	400	53.3	0.52	39.00	4.338	74.4
289	400	400	53.0	0.53	39.00	1.971	74.5
305	400	400	53.0	0.53	39.00	1.971	74.5
309	400	400	54.0	0.53	38.65	1.957	74.7
314	400	400	53.5	0.53	38.75	1.968	74.6

[1] Normalized to 400 psi - 68°F

TABLE I-1 (Continued)

HOLLOW FIBER OPERATING DATA
FOR SINGLE PERMEATOR

Elapsed Time (hrs)	Pressure In (psi)	Pressure Out (psi)	Temp. °F	Brine Flow gpm	Product Flow sec/gal	Normalized[1] Product Flow gpm	Product Water Recovery - %
329	400	400	53.0	0.53	39.50	1.956	74.3
337	400	400	54.0	0.53	39.0	1.939	74.5
355	400	400	54.0	0.52	39.40	1.919	74.4
377	400	400	52.5	0.52	40.10	1.933	74.2
378	400	400	54.0	0.55	37.30	2.017	74.3
381	400	400	53.0	0.55	38.50	1.997	73.8
394	400	400	52.5	0.55	39.00	1.988	73.7
405	400	400	52.5	0.54	38.50	2.014	74.1
419	400	400	52.5	0.54	38.50	2.011	74.3
422	400	400	53.0	0.53	38.50	1.993	74.5
432	400	400	52.0	0.53	38.90	1.978	74.5
443	400	400	52.5	0.53	38.90	1.993	74.6
449	400	400	52.5	0.52	38.40	1.986	74.9
478	400	400	53.0	0.49	39.20	1.961	74.7
478	400	400	53.0	0.54	39.20	1.961	73.8
491	400	400	53.0	0.54	38.50	1.997	74.2
523	405	400	53.0	0.53	39.60	1.930	73.9
527	405	400	53.0	0.53	38.90	1.964	64.6
539	400	390	53.0	0.51	40.30	1.932	74.5
546	390	385	53.5	0.50	41.00	1.920	74.6
563	405	400	52.9	0.51	39.50	1.938	74.7
575	405	400	53.0	0.52	39.65	1.927	74.6
587	405	398	53.0	0.51	39.70	1.929	74.7
595	405	400	54.0	0.52	39.60	1.898	74.6
615	415	400	53.0	0.52	39.50	1.911	74.6
636	405	390	54.0	0.51	40.40	1.884	74.4
641	390	385	54.0	0.50	41.50	1.881	74.4
660	402	390	53.5	0.51	40.10	1.921	74.6
685	410	390	53.5	0.50	41.00	1.860	74.4
693	400	385	54.2	0.52	39.30	1.955	74.7
698	400	387	55.0	0.51	39.90	1.895	74.7
705	400	390	54.5	0.47	40.10	1.894	76.1
718	400	383	54.0	0.46	40.35	1.915	76.3
718	400	383	54.0	0.47	40.40	1.913	75.8
740	400	380	54.0	0.52	41.40	1.874	73.4
764	400	380	54.0	0.51	41.10	1.887	74.2
772	400	380	55.0	0.48	40.85	1.868	75.5
789	400	383	54.8	0.48	40.20	1.897	75.6
812	400	380	55.0	0.46	43.90	1.738	74.6
838	400	380	54.8	0.47	42.00	1.823	75.1

[1] Normalized to 400 psi - 68°F

TABLE I-2

HOLLOW FIBER OPERATING DATA 2-1 ARRAY

Elapsed Time hrs.	Pressure, psi In	2nd stage	Out	Feed Temp °F	Brine Flow sec/gal	Product Water Flow (sec/gal Module 1129	1131	691
1	400	345	315	54.6	39.6	28.4	28.9	50.0
9	400	345	315	53.0	39.6	28.7	29.5	51.0
22	400	350	315	54.5	40.7	28.6	29.2	49.9
70	400	350	315	54.5	42.7	28.7	29.1	49.1
97	400	345	310	54.0	38.4	29.5	30.1	51.3
116	410	345	305	53.0	36.8	29.8	30.3	52.5
140	410	349	305	53.0	37.3	29.9	30.5	52.0
140	410	340	300	53.0	35.5	30.2	30.7	53.0
145	410	347	304	54.0	35.5	29.7	30.2	52.5
163	410	342	300	53.0	35.7	30.0	30.4	52.5
187	410	340	300	54.0	35.3	30.2	30.9	52.6
191	410	345	304	54.7	35.0	29.5	29.8	51.2
212	410	348	305	54.0	36.0	29.8	30.3	51.5
235	410	348	303	53.5	36.0	29.9	30.4	52.0
259	410	340	302	54.0	35.4	30.0	30.7	51.6
284	410	342	300	54.0	35.5	30.1	30.6	52.1
308	410	347	305	54.0	35.5	29.9	30.3	51.3
332	420	345	302	55.0	35.6	29.8	30.4	51.5
340	410	345	308	54.0	36.1	29.8	30.5	51.1
355	410	345	301	54.0	35.7	30.2	30.8	51.8
359	400	333	293	54.5	37.5	30.5	31.3	52.6
380	400	330	290	54.0	37.8	31.2	32.0	53.5
386	400	350	320	54.5	69.0	30.5	31.2	49.5
390	400	347	320	53.8	68.3	31.1	31.8	50.3
404	400	347	317	53.8	68.1	31.0	31.4	50.6
414	400	346	316	54.0	68.4	31.1	32.0	51.1
426	400	346	316	53.5	68.6	31.6	32.1	52.3
431	400	330	292	54.0	40.5	31.5	32.0	55.0
432	400	327	289	54.5	39.0	31.0	32.2	55.9
447	400	325	285	54.5	39.3	30.9	31.5	55.4
452	400	325	285	54.5	38.7	30.7	31.5	54.8
471	400	300	290	55.0	40.8	30.9	31.5	54.2
476	400	325	285	55.0	40.2	30.7	31.6	54.8
484	400	325	285	54.0	40.8	30.9	31.6	55.0
497	400	330	290	54.5	52.2	30.6	30.7	54.0
516	400	330	290	54.0	40.5	30.5	30.8	54.2
540	400	330	290	54.0	40.1	30.5	30.7	53.3
562	410	340	300	54.0	40.3	30.6	30.6	55.0
568	400	328	290	54.5	41.9	31.4	31.9	55.6

TABLE I-2 (Continued)

HOLLOW FIBER OPERATING DATA 2-1 ARRAY

Elapsed Time hrs.	Pressure, psi In	2nd stage	Out	Feed Temp °F	Brine Flow sec/gal	Product Water Flow (sec/gal) Module 1129	1131	691
578	400	340	320	54.0	28.0	31.3	31.7	49.5
595	400	338	312	54.0	57.6	31.9	32.7	53.3
499	400	431	319	54.0	67.7	31.9	32.7	53.1
512	400	340	318	54.0	67.9	32.0	32.5	53.6
620	405	342	320	54.0	69.4	31.8	32.5	54.0
636	400	340	315	54.0	69.5	32.2	32.9	55.2
646	400	338	313	54.0	70.2	32.7	33.3	56.9
661	400	342	318	54.5	70.2	32.1	32.7	56.8
667	303	343	319	54.0	69.8	32.4	32.8	57.9
669	400	334	309	54.0	52.6	32.4	33.0	58.5
684	400	334	309	54.5	52.6	32.5	33.3	59.9
692	400	334	309	54.0	52.5	32.4	33.1	59.9
706	400	331	301	54.0	52.5	32.8	33.4	60.8
713	400	322	290	54.0	41.4	33.1	33.8	63.2
729	400	321	290	53.0	41.8	33.4	34.1	69.9
741	400	325	295	54.0	42.6	33.3	33.6	62.4
756	400	325	292	54.0	42.6	33.0	33.7	62.6
760	400	335	305	54.2	49.0	32.6	33.1	60.3
778	400	335	303	54.0	49.7	32.9	33.2	61.1
802	400	332	303	54.0	51.4	33.1	33.5	62.1
829	400	332	302	54.0	51.7	32.9	33.4	62.5
852	400	330	302	54.8	52.0	33.1	33.4	63.2
906	400	330	300	54.8	52.5	32.9	33.5	64.0
922	400	333	302	54.0	53.5	33.2	33.7	65.4
947	400	335	302	54.0	53.4	32.8	33.4	66.2
972	410	343	310	54.2	53.0	32.7	32.9	65.4
975	401	330	310	55.0	47.5	32.1	32.4	61.6
978	400	325	310	54.0	46.9	32.6	33.7	61.3
987	400	328	310	54.0	46.3	32.7	33.1	61.6
1011	400	328	320	54.5	46.1	32.6	33.7	62.5
1016	400	328	320	55.0	45.8	32.7	33.4	62.5
1019	400	320	301	55.0	46.3	32.5	33.4	61.9
1023	400	321	309	55.0	46.4	32.7	33.4	62.5
1033	400	322	309	54.3	46.4	32.5	33.0	62.5
1059	400	326	310	56.0	40.9	32.2	32.9	61.3
1087	400	326	310	55.0	45.8	32.4	33.6	63.0
1111	400	325	310	55.0	46.3	32.8	33.4	63.3
1131	400	325	310	55.0	46.3	32.6	33.4	63.3
1150	400	325	310	55.0	46.0	32.8	33.6	63.1
1171	400	325	310	55.0	46.6	32.6	33.4	63.6
1195	400	325	310	55.0	46.5	32.9	33.5	64.1
1221	400	325	310	55.0	46.5	32.7	33.4	64.2

TABLE I-2 (Continued)

HOLLOW FIBER OPERATING DATA 2-1 ARRAY

Elapsed Time Hrs.	Pressure, psi In	2nd stage	Out	Feed Temp °F	Brine Flow sec/gal	Product Water Flow (sec/gal) Module 1129	1131	691
1226	400	331	311	55.1	46.8	32.4	33.2	63.8
1243	410	335	315	54.5	47.0	32.7	33.7	64.2
1256	400	320	305	54.5	40.4	33.4	33.5	59.6
1249	400	320	305	54.0	40.6	33.5	33.9	60.5
1273	405	333	320	54.0	51.4	32.6	33.4	56.5
1289	405	325	310	52.5	39.2	33.6	34.0	60.5
1315	400	320	300	53.5	39.8	33.9	34.2	61.1
1341	400	320	302	54.8	39.5	33.8	34.2	60.4
1366	402	320	305	55.1	39.8	33.6	34.2	60.1
1386	400	315	300	54.0	40.1	34.6	35.0	61.7
1395	402	320	305	55.0	40.3	33.7	34.2	60.5
1418	400	320	305	55.0	43.2	34.1	34.6	60.8
1442	400	318	300	55.0	41.8	34.4	34.8	61.2
1462	400	318	300	54.0	41.6	34.9	35.5	62.8
1484	400	318	300	54.0	40.0	34.7	35.3	62.2
1532	400	315	298	54.0	40.4	35.5	35.7	63.0
1561	400	315	298	55.0	40.7	35.6	35.9	63.2
1580	400	315	298	54.0	40.5	35.3	35.7	63.6
1603	400	315	298	54.0	40.6	35.3	35.8	63.4
1628	400	316	297	54.0	40.8	35.5	35.8	63.7
1642	400	315	298	54.5	41.2	34.5	34.9	61.8
1689	400	315	298	54.5	41.4	34.7	35.1	62.3
1718	400	308	295	55.0	41.2	34.9	35.3	63.0
1738	400	310	296	54.5	41.1	35.1	35.8	63.8
1764	400	310	295	55.0	41.1	35.1	35.7	62.8
1786	400	308	292	55.0	41.6	35.4	35.7	63.0
1814	400	305	290	54.0	41.3	35.5	35.8	63.4
1815	400	315	298	54.0	41.0	33.2	34.0	59.6
1825	400	320	300	54.0	40.8	33.3	33.6	59.0

TABLE I - 3

OPERATIONAL DATA HOLLOW FIBER RO SYSTEM
(2-1 ARRAY FIRST STAGE MODULES)

Elapsed Time (hrs)	Normalized Product Water Flow gpm Module 1129	Normalized Product Water Flow gpm Module 1131	Brine Flow gpm Each Module	Water Recovery % Each Module	Bundle Pressure Drop psi
1	2.679	2.637	1.36	60.6	14
9	2.729	2.655	1.35	60.5	14
22	2.652	2.602	1.34	60.8	10
70	2.652	1.616	1.31	61.2	11
97	2.608	2.556	1.35	59.6	13
116	2.587	2.545	1.39	59.0	22
140	2.567	2.516	1.38	58.9	18
140	2.564	2.523	1.41	58.2	25
145	2.541	2.499	1.42	58.5	18
163	2.575	2.541	1.41	58.4	23
187	2.520	2.463	1.42	58.0	25
191	2.530	2.505	1.44	58.3	18
212	2.530	2.488	1.42	58.5	17
235	2.543	2.501	1.41	58.5	17
259	2.535	2.474	1.43	58.0	24
284	2.522	2.477	1.42	58.1	23
308	2.522	2.484	1.43	58.2	17
332	2.465	2.416	1.43	58.3	29
340	2.534	2.480	1.42	58.3	20
355	2.505	2.452	1.42	58.0	20
359	2.539	2.474	1.37	58.6	25
380	2.516	2.453	1.35	58.3	29
386	2.541	2.484	1.04	65.1	25
390	2.532	2.476	1.04	64.8	29
404	2.541	2.508	1.03	65.0	29
414	2.528	2.457	1.03	64.9	30
426	2.511	2.472	1.01	65.0	31
431	2.505	2.466	1.29	59.5	33
432	2.482	2.436	1.31	59.0	35
447	2.546	2.497	1.30	59.5	37
452	2.559	2.494	1.32	59.3	37
471	2.613	2.564	1.29	59.8	63
476	2.543	2.471	1.29	59.8	37
484	2.571	2.514	1.28	59.9	38
497	2.586	2.577	1.13	63.3	41
516	2.585	2.560	1.29	60.2	32
540	2.582	2.565	1.31	59.9	31
562	2.512	2.496	1.29	60.2	33
568	2.504	2.565	1.26	60.1	37

TABLE I - 3 (Continued)

OPERATIONAL DATA HOLLOW FIBER RO SYSTEM
(2-1 ARRAY FIRST STAGE MODULES)

Elapsed Time (hrs)	Normalized Product Water Flow gpm Module 1129	Module 1131	Brine Flow gpm Each Module	Water Recovery % Each Module	Bundle Pressure Drop psi
578	2.408	2.378	1.68	53.1	35
595	2.482	2.421	1.08	6.31	35
599	2.222	2.168	1.01	64.8	35
612	2.480	2.442	1.00	65.0	37
620	2.471	2.422	0.99	65.3	41
636	2.469	2.416	0.98	65.4	38
646	2.440	2.396	9.95	65.5	41
661	2.448	2.403	0.96	65.9	37
667	2.435	2.405	0.95	66.0	41
669	2.457	2.412	1.08	62.8	39
684	2.431	2.372	1.07	63.0	40
692	2.455	2.406	1.07	63.0	40
706	2.439	2.392	1.06	62.9	43
713	2.424	2.374	1.20	59.9	46
729	2.455	2.401	1.15	60.7	49
741	2.403	2.381	1.18	60.2	43
756	2.425	2.374	1.18	60.3	43
760	2.426	2.390	1.11	62.2	37
778	2.414	2.393	1.09	62.3	38
802	2.414	2.385	1.07	63.8	42
829	2.429	2.393	1.06	63.0	43
852	2.390	2.369	1.05	63.1	45
906	2.407	2.364	1.04	63.4	45
922	2.410	2.374	1.02	63.7	43
947	2.434	2.390	1.01	64.1	42
971	2.375	2.361	1.02	64.0	43
975	2.443	2.417	1.12	62.4	43
978	2.464	2.383	1.13	61.5	46
987	2.445	2.416	1.13	61.6	43
1011	2.430	2.354	1.13	61.5	43
1016	2.405	2.351	1.14	61.5	43
1019	2.446	2.381	1.13	61.6	51
1023	2.429	2.378	1.13		
1033	2.469	2.432	1.13	61.9	49
1059	2.394	2.343	1.22	60.1	40
1087	2.434	2.348	1.13	61.6	45
1111	2.409	2.366	1.12	61.7	47
1131	2.424	2.366	1.12	61.8	47
1150	2.409	2.451	1.13	61.5	46
1171	2.425	2.367	1.12	61.9	47
1195	2.404	2.360	1.11	61.8	47
1221	2.418	2.368	1.11	62.0	47

TABLE I - 3 (Continued)

OPERATIONAL DATA HOLLOW FIBER RO SYSTEM
(2-1 ARRAY FIRST STAGE MODULES)

Elapsed Time (hrs)	Normalized Product Water Flow gpm Module 1129	Module 1131	Brine Flow gpm Each Module	Water Recovery % Each Module	Bundle Pressure Drop psi
1226	2.418	2.359	1.11	62.2	41
1243	2.376	2.306	1.11	62.0	47
1256	2.381	2.370	1.25	59.0	45
1249	2.396	2.367	1.23	59.0	46
1273	2.424	2.366	1.11	61.9	44
1289	2.412	2.383	1.26	58.4	44
1315	2.385	2.361	1.24	58.5	45
1341	2.339	2.312	1.26	58.4	44
1366	2.336	2.295	1.25	58.5	47
1386	2.335	2.308	1.23	58.2	51
1395	2.335	2.301	1.24	58.7	47
1418	2.323	2.289	1.19	59.5	48
1442	2.305	2.279	1.21	58.9	49
1462	2.312	2.272	1.20	58.7	50
1484	2.319	2.280	1.23	58.1	48
1532	2.278	2.266	1.22	58.0	52
1561	2.236	2.217	1.21	58.0	52
1580	2.279	2.267	1.21	58.1	52
1603	2.292	2.260	1.21	58.2	52
1628	2.277	2.258	1.21	58.2	51
1642	2.326	2.299	1.21	58.7	52
1689	2.314	2.287	1.21	58.7	52
1718	2.303	2.277	1.20	58.6	59
1738	2.304	2.259	1.20	58.5	58
1764	2.284	2.245	1.21	58.3	57
1786	2.272	2.253	1.20	58.4	60
1814	2.312	2.293	1.20	58.3	63
1815	2.433	2.376	1.24	59.1	51
1825	2.408	2.387	1.24	59.0	45
1832	2.396	2.353	1.24	59.0	45

TABLE I - 4

HOLLOW FIBER SYSTEM OPERATION
(2-1 ARRAY, 2ND STAGE MODULE)

Elapsed Time (Hrs)	Normalized Product Water Flow gpm Module 691	Brine Flow Rate gpm	Overall System Recovery %	Bundle Pressure Drop psi
839	1.815	1.51	78.1	30
847	1.827	1.52	77.8	30
860	1.808	1.47	78.4	35
908	1.838	1.41	79.3	35
935	1.801	1.56	76.9	35
954	1.802	1.63	75.9	40
978	1.809	1.61	76.1	44
979	1.813	1.69	75.0	40
983	1.770	1.69	75.3	43
1002	1.825	1.68	75.3	42
1026	1.797	1.70	74.9	40
1030	1.797	1.71	75.3	40
1050	1.799	1.67	75.6	43
1073	1.802	1.67	75.5	45
1098	1.826	1.69	75.1	38
1122	1.809	1.69	75.1	42
1146	1.809	1.69	75.3	42
1170	1.786	1.69	75.4	43
1178	1.813	1.66	75.6	37
1193	1.808	1.68	75.2	44
1197	1.822	1.60	75.8	40
1218	1.824	1.59	75.6	40
1224	1.809	0.87	85.4	30
1228	1.809	0.88	85.1	27
1242	1.807	0.88	85.1	30
1252	1.788	0.88	85.0	30
1264	1.762	0.87	84.9	30
1269	1.769	1.48	76.7	38
1271	1.743	1.54	75.9	38
1285	1.776	1.53	76.4	40
1290	1.795	1.55	76.2	40
1309	1.861	1.47	77.1	10
1315	1.780	1.49	76.8	40
1322	1.803	1.47	77.0	40
1335	1.792	1.15	81.4	40
1355	1.800	1.48	77.2	40
1378	1.831	1.50	77.1	40
1401	1.719	1.49	77.1	40
1406	1.745	1.43	77.3	38
1416	1.852	2.14	70.1	30

TABLE I - 4 (Continued)

HOLLOW FIBER SYSTEM OPERATION
(2-1 ARRAY, 2ND STAGE MODULE)

Elapsed Time (Hrs)	Normalized Product Water Flow gpm Module 691	Brine Flow Rate gpm	Overall System Recovery %	Bundle Pressure Drop psi
1434	1.746	1.04	82.3	26
1437	1.518	0.89	84.5	11
1450	1.715	0.88	84.6	22
1459	1.692	0.86	84.8	22
1474	1.673	0.86	84.7	25
1484	1.633	0.85	84.6	25
1499	1.601	0.85	84.8	24
1506	1.578	0.86	84.6	24
1507	1.608	1.14	80.5	25
1522	1.558	1.14	80.3	25
1530	1.571	1.14	80.3	25
1545	1.574	1.14	80.1	30
1551	1.564	1.45	75.8	32
1567	1.440	1.44	75.5	31
1580	1.564	1.41	76.4	30
1594	1.566	1.41	76.4	33
1599	1.563	1.22	79.1	30
1616	1.552	1.21	79.3	32
1640	1.534	1.17	79.7	29
1667	1.527	1.16	79.8	30
1691	1.495	1.15	79.8	28
1744	1.481	1.14	79.9	30
1761	1.457	1.12	80.1	31
1785	1.435	1.12	80.1	33
1810	1.412	1.13	80.2	33
1813	1.509	1.26	78.8	20
1816	1.554	1.28	78.2	15
1825	1.539	1.30	78.1	18
1850	1.482	1.30	77.9	8
1855	1.469	1.31	77.8	8
1857	1.548	1.29	78.1	19
1862	1.511	1.29	78.0	12
1872	1.527	1.29	78.1	13
1897	1.501	1.47	76.1	16
1925	1.485	1.31	77.8	16
1949	1.480	1.30	77.9	15
1969	1.480	1.30	78.0	15
1988	1.485	1.30	77.8	15
2010	1.474	1.29	78.1	15
2034	1.462	1.29	77.9	15

TABLE I - 4 (Continued)

HOLLOW FIBER SYSTEM OPERATION
(2-1 ARRAY, 2ND STAGE MODULE)

Elapsed Time (Hrs)	Normalized Product Water Flow gpm Module 691	Brine Flow Rate gpm	Overall System Recovery %	Bundle Pressure Drop psi
2059	1.460	1.29	78.0	15
2065	1.450	1.28	78.2	20
2081	1.438	1.28	78.1	20
2094	1.609	1.49	75.6	15
2087	1.600	1.48	75.5	15
2112	1.640	1.17	80.1	13
2128	1.614	1.53	74.8	15
2153	1.610	1.51	74.9	20
2179	1.589	1.52	74.9	18
2204	1.582	1.51	75.1	15
2225	1.594	1.50	74.7	15
2233	1.574	1.49	75.2	15
2256	1.566	1.39	76.3	15
2280	1.573	1.44	75.6	18
2300	1.559	1.44	75.2	18
2322	1.574	1.50	74.5	18
2371	1.567	1.49	74.4	17
2399	1.536	1.47	74.5	17
2418	1.552	1.48	74.4	17
2442	1.557	1.48	74.5	17
2466	1.549	1.47	74.6	19
2480	1.584	1.46	75.3	17
2527	1.571	1.45	75.2	17
2556	1.566	1.46	75.0	13
2576	1.552	1.46	74.8	14
2602	1.566	1.46	74.9	15
2624	1.574	1.44	75.0	16
2652	1.604	1.45	74.8	15
2653	1.656	1.46	75.8	17
2663	1.654	1.47	75.8	20
2670	1.646	1.46	75.9	18

TABLE 1 - 5

ANALYSIS DATA HOLLOW FIBER SYSTEM (AA DATA)

Hour Clock RDG Hrs.	Ca mg/l	Mg mg/l	Mn mg/l	Total Iron mg/l	Silica mg/l	Al mg/l	TDS mg/l
			FEED WATER				
1234	115	85	13.8	71	11	8.1	1281
1302	115	86	14.3	74	10	7.8	1340
1395	115	87	14.5	75	10	8.0	1337
1590	97	74	12.4	54	10	7.4	1044
1899	97	72	12.7	57	10	8.3	-
2051	102	77	13.2	59	11	7.9	1248
2314	107	77	13.5	69	11	8.5	1278
2528	112	78	14.8	74	11	8.5	1277
2904	120	89	14.6	80	11	8.4	1498
3074	125	94	15.0	84	12	9.0	1510
3162	123	90	14.9	80	11	8.6	1387
3489	103	82	13.8	64	12	9.0	1313
			FINAL BRINE				
1234	480	378	58.3	315	45	31.2	5567
1302	452	360	54.6	298	42	26.0	5165
1395	422	335	52.0	284	40	31.0	4791
1590	376	312	47.2	212	42	29.0	4153
1899	384	280	25.2	223	43	32.0	-
2051	404	307	51.0	241	49	35.4	4862
2314	433	320	53.3	292	43	34.0	5235
2528	715	572	86.4	493	76	54.0	8197
2904	576	436	67.0	386	53	39.0	7154
3074	624	495	71.7	426	57	40.0	7394
3162	568	438	67.2	376	51	35.0	6389
3489	407	337	50.0	254	49	29.0	4994

TABLE I-6

ANALYSIS DATA HOLLOW FIBER SYSTEM (AA DATA)

Hour Clock Rdg Hrs	Ca mg/l	Mg mg/l	Mn mg/l	Total Iron mg/l	Silica mg/l	Al mg/l	TDS mg/l
			FIRST STAGE BRINE				
2314	257	184	32.4	167	26	20	3104
2528	316	225	43.2	212	31	23	3618
2904	319	233	36.8	205	30	22	4035
3074	339	268	41.3	236	31	24	4188
3162	328	242	38.8	211	29	22	3704
3489	253	201	31.8	154	27	21	3157
			PRODUCT WATER PERMEATOR #1129				
2314	0.71	0.54	0.10	0.79	0.8	0.1	24
2528	0.55	0.44	0.04	0.76	1.0	0.1	14
2904	0.73	0.49	0.08	0.43	1.1	0.4	28
3074	0.73	0.55	0.14	0.45	2.0	0.4	-
3162	0.78	0.58	0.09	0.51	1.0	0.1	16
3489	0.58	0.43	0.09	0.34	1.0	0.1	27
			PRODUCT WATER PERMEATOR #1131				
2314	0.85	0.66	0.12	1.04	0.8	0.1	26
2528	0.70	0.56	0.06	0.96	1.0	0.1	11
2904	0.77	0.51	0.09	0.43	1.3	0.4	27
3074	0.66	0.52	0.13	0.42	2.0	0.3	-
3162	0.69	0.55	0.09	0.47	1.0	0.3	41
3489	0.54	0.44	0.09	0.36	1.0	0.1	31
			PRODUCT WATER PERMEATOR #691				
1234	0.40	0.44	0.05	0.27	0.50	0.10	14
1302	0.52	0.92	0.15	0.70	0.40	0.40	4
1395	0.51	0.72	0.10	0.49	0.75	0.18	10
1590	0.60	0.50	0.05	0.40	0.50	0.20	33
1899	0.56	0.40	0.08	0.32	0.40	0.04	-
2051	0.74	0.56	0.08	0.44	0.40	0.15	45
2314	1.06	0.77	0.15	1.19	0.8	0.10	34
2528	1.09	1.23	0.20	1.82	1.0	0.10	9
2904	2.21	1.47	0.25	0.27	1.3	0.40	40
3074	2.68	2.07	0.42	2.15	2.0	0.50	-
3162	2.09	1.48	0.25	1.42	1.0	0.40	29
3489	1.48	1.18	0.22	0.99	1.0	0.10	30

TABLE I-7

FIELD ANALYSIS DATA HOLLOW FIBER SYSTEM

Hour Clock Rdg	Total Iron mg/l	Iron II mg/l	SO_4 mg/l	Total Hardness ($CaCO_3$) mg/l	Calcium Hardness ($CaCO_3$) mg/l	pH Units	Meter TDS mg/l
			RAW AMD				
1253	74	66	890	560	320	3.3	-
1282	67	72	800	580	328	3.5	-
1301	74	67	730	728	324	3.4	-
1328	75	66	860	716	340	-	-
1348	65	65	890	684	308	3.6	1300
1380	71	70	725	600	380	3.6	-
1445	76	70	760	640	280	3.3	1275
1632	54	49	710	480	240	3.6	1200
1663	50	50	580	500	225	3.5	-
1682	56	47	620	500	250	3.3	-
1778	55	51	590	500	275	3.5	1300
1826	50	50	700	350	340	3.4	-
1938	55	51	890	465	305	-	-
1956	56	56	720	588	255	-	1250
2051	56	56	690	550	300	-	-
2314	57	57	780	560	290	-	1100
2366	68	65	760	580	300	-	1100
2528	70	69	663	650	300	3.3	1100
2717	100	76	710	650	250	3.4	1200
2810	88	78	920	650	300	3.4	1200
3031	66	73	780	720	290	-	1400
3051	70	76	950	700	310	-	1350
3114	75	75	890	800	250	-	1250
3370	78	78	940	680	320	-	1400
			INTER-STAGE BRINE				
2314	150	142	1300	1400	600	-	2500
2366	149	118	1500	1650	650	-	2300
2482	155	150	1700	1400	600	3.0	2500
2528	190	190	1950	1600	800	2.9	2650
2717	240	185	1700	1500	750	3.1	2900
2810	210	190	2400	1600	800	3.2	2900
3031	186	210	2600	1680	860	-	3450
3051	235	185	2350	1800	680	-	3400
3114	195	198	2350	2200	1000	-	3100
3370	190	181	2000	1700	700	-	3200

TABLE I - 8

FIELD ANALYSIS DATA HOLLOW FIBER SYSTEM

Hour Clock RDG Hrs.	Total Iron mg/l	Iron II mg/l	SO_4 mg/l	Tota Hardness Ca CO_3 mg/l	Calcium Hardness Ca CO_3 mg/l	pH Units	Meter TDS mg/l
			BRINE				
1253	304	305	3600	3040	1400	2.8	4750
1282	275	270	2800	2000	1740	3.0	4350
1301	285	300	2700	2420	1440	2.8	4600
1328	388	225	2500	3100	1640	-	4300
1348	300	260	2800	2520	1200	-	3900
1380	315	290	3100	2300	1300	2.9	4100
1445	255	208	3150	2400	1300	3.1	3950
1632	189	182	2500	1900	900	3.0	3900
1663	189	183	2450	2000	850	2.6	3900
1682	199	182	2500	2100	1000	2.7	3900
1778	200	175	2500	2070	920	2.8	3950
1826	150	167	2450	2070	940	2.9	3850
1938	220	260	2600	2200	1020	-	3800
1956	220	200	3000	2300	1010	-	3750
2051	230	230	2040	2250	1990	-	3800
2314	260	270	2600	2200	1000	-	4000
2366	305	290	2500	2200	1100	-	3600
2482	280	280	2800	2600	1100	2.9	4000
2528	498	464	6000	4100	1700	2.7	7800
2717	540	510	5000	4000	1500	2.9	9000
2810	450	395	4500	3250	1500	2.9	4800
3031	370	415	3800	3500	1650	-	-
3051	450	255	3800	3600	1500	-	8500
3114	390	345	3850	3100	1400	-	4950
3370	340	328	3100	2600	1400	-	4900
			PRODUCT WATER PERMEATOR 1129				
2314	0.47	0.42	2	5.0	5.0	-	12
2366	0.48	0.41	5	5.5	1.0	-	12
2482	0.45	0.45	2	3.0	1.0	3.9	11
2528	0.45	0.40	2	4.0	2.0	4.3	13
2717	0.51	0.50	1	4.0	2.0	4.4	14
2810	0.58	0.57	1	4.0	2.0	4.5	12
3031	0.49	0.49	4	3.8	3.0	-	10
3051	0.48	0.48	2	4.0	3.6	-	11
3114	0.50	0.49	2	5.0	2.0	-	12
3370	0.55	0.52	2	3.0	1.0	-	11

TABLE I - 9

FIELD ANALYSIS DATA HOLLOW FIBER SYSTEM

Hour Clock RDG Hrs.	Total Iron mg/l	Iron II mg/l	SO_4 mg/l	Total Hardness Ca CO_3 mg/l	Calcium Hardness Ca CO_3 mg/l	pH Units	Meter TDS mg/l
			PRODUCT WATER PERMEATOR 691				
1253	0.81	0.76	3	-	-	4.5	18
1282	0.77	0.77	4	-	-	4.5	18
1301	0.731	0.70	3	-	-	4.4	16
1328	0.79	0.79	5	-	-	-	17
1348	0.70	0.65	3	-	-	4.6	14
1380	0.60	0.58	1	2.5	2.5	4.8	14
1445	0.62	0.52	1	4.0	2.0	4.6	14
1632	0.41	0.38	1	2.5	2.5	4.8	10
1663	0.38	0.34	1	2.0	1.0	4.4	10
1682	0.35	0.32	1	2.0	1.0	4.4	11
1778	0.36	0.33	1	10.8	2.0	4.5	12
1826	0.33	0.32	2	0.7	0.2	4.5	12
1938	0.59	0.54	2	5.5	4.5	-	11
1956	0.46	0.45	3	5.5	4.5	-	11
2051	0.45	0.42	2	5.5	4.0	-	13
2314	0.55	0.53	4	13.0	1.5	-	17
2366	0.61	0.51	3	13.0	5.0	-	16
2482	0.60	0.59	3	5.0	2.0	3.6	17
2528	1.10	1.10	3	8.0	3.0	4.0	25
2717	1.34	1.31	8	10.0	3.0	4.1	26
2810	1.20	1.20	10	10.0	4.0	4.3	25
3031	1.59	1.60	17	12.6	5.8	-	28
3051	1.61	1.68	19	14.0	6.8	-	29
3114	1.40	1.36	12	12.0	4.7	-	28
3370	1.55	1.52	10	10.0	4.0	-	26
			PRODUCT WATER PERMEATOR 1131				
2314	0.56	0.52	2	5.0	3.0	-	12
2366	0.59	0.55	3	5.5	4.0	-	11
2482	0.50	0.50	2	3.0	1.0	3.7	11
2528	0.52	0.50	1	4.0	2.0	4.2	13
2717	0.52	0.50	1	4.0	2.0	4.4	15
2810	0.55	0.55	1	4.0	2.0	4.5	12
3031	0.50	0.47	4	3.2	1.2	-	10
3051	0.52	0.50	3	3.0	1.0	-	11
3114	0.53	0.51	1	7.0	2.0	-	12
3370	0.49	0.48	3	3.0	1.0	-	11

II. Tubular RO System Data

This section contains the data collected on the tubular RO system.

The following equations were utilized for normalizing the flux data.

For Temperature:

$$Q_{68} = Q_o * (1.01667^{(68-T)})$$

where: Q_{68} = Product flow @ 68°F (gpm)

Q_o = Product flow at observed temperature (gpm)

T = Temperature observed

For Pressure:

$$Q_{600} = Q_o * (1200/P_1 + P_2)$$

where: Q_{600} = Product flow at 600 psi

Q_o = Observed flow gpm

P_1 = Pressure in

P_2 = Pressure out

Therefore Normalized Flux is:

$$F_{600-68} = Q_o(1200/P_1 + P_2) \times (1.01667^{(68-T)}) \times (1440/N \times 16.9)$$

where: F_{600-68} = Flux at 600 psi - 68°F in gafd

N = Number of modules utilized

NOTE: 1440 converts gpm to gpd & 16.9 sq ft membrane/module

TABLE II - 1

TUBULAR RO SYSTEM OPERATIONAL DATA 310 MODULES

Elapsed Time (hrs)	Pressure In psi	Pressure Out psi	Temp °F	Brine Rate sec/gal	Product Rate sec/5 gal	Flux[1] gsfd	Product Water Recovery %
Phase I - 60 Modules							
23	600	410	52.7	31.7	50.5	12.9	75.8
41	600	400	51.8	26.8	53.7	12.5	71.4
64	600	425	51.8	33.3	53.9	12.1	75.5
80	605	420	52.5	31.3	53.4	12.1	74.4
96	600	422	53.2	33.2	55.6	11.5	74.9
112	602	422	52.3	32.2	57.7	11.2	73.6
125	600	432	53.8	35.2	58.8	10.7	74.9
145	600	320	53.0	33.4	60.6	10.5	73.4
161	601	432	52.9	33.9	60.9	10.4	73.6
174	600	389	53.0	27.9	63.6	10.5	68.3
188	600	415	53.0	30.8	59.5	10.9	72.1
194	600	411	52.0	21.9	57.7	11.1	74.0
207	610	430	52.9	33.3	57.0	11.1	74.5
218	600	420	52.9	31.9	58.9	10.9	73.0
240	600	420	53.5	31.0	59.0	10.8	72.4
255	600	400	53.0	30.4	60.2	10.9	61.6
280	600	400	53.0	28.3	61.5	10.7	69.7
304	600	395	53.0	25.4	64.0	10.4	66.5
328	600	355	53.0	23.4	65.6	10.5	64.1
336	600	370	54.0	24.5	63.7	10.4	65.7
381	605	330	53.0	21.3	70.9	9.9	60.0
407	620	339	53.5	20.9	72.0	9.4	59.0
430	620	300	54.0	18.8	77.46	9.2	55.0
448	625	302	54.0	18.64	77.40	9.0	54.6
470	630	305	52.5	18.4	81.50	8.7	53.0
Phase II - 35 Modules							
492	700	485	52.5	32.5	92.7	10.3	63.8
503	700	470	52.5	28.8	98.4	9.8	59.4
517	700	470	52.5	28.2	104.3	9.3	57.5
529	700	455	52.0	26.1	108.1	9.1	54.7
541	700	440	52.5	24.8	113.0	8.8	52.4
546	700	375	53.5	20.9	116.2	8.9	47.4
575	700	440	53.0	24.7	118.5	8.3	51.0
589	700	440	53.0	24.3	120.3	8.2	50.2
620	700	432	53.0	23.6	127.8	7.8	48.0
625	700	432	53.0	23.3	129.2	7.7	47.4
661	700	459	52.9	25.3	131.3	7.4	49.0

[1] Normalized to 68°F - 600 psi

TABLE II - 1 (Continued)

TUBULAR RO SYSTEM OPERATIONAL DATA 310 MODULES

Phase II - 35 Modules

Elapsed Time (hrs)	Pressure In psi	Pressure Out psi	Temp °F	Brine Rate sec/gal	Product Rate sec/5 gal	Flux[1] gsfd	Product Water Recovery %
673	705	443	53.0	23.2	133.5	7.2	46.5
685	700	447	53.0	24.1	134.4	7.3	47.3
712	698	400	53.0	21.44	142.5	7.2	42.9
734	700	450	54.0	24.5	137.9	7.0	47.0
757	700	500	53.5	29.2	134.8	6.9	52.0
782	700	445	53.5	24.0	141.5	6.9	45.9
803	700	455	54.5	25.1	121.5	7.8	50.8
815	700	455	54.0	25.0	128.1	7.5	49.4
838	700	462	54.0	24.6	128.4	7.4	48.9
849	700	440	53.5	23.7	130.0	7.5	47.7
862	700	470	54.0	25.3	133.4	7.1	48.6
887	699	438	54.8	23.5	133.9	7.2	46.7
935	702	420	54.8	22.1	138.7	7.0	44.3

Phase III & IV - 15 Modules

					(sec/gal)		
961	570	452	53.0	99.5	71.1	7.2	58.3
971	570	460	53.0	98.3	71.5	7.1	57.9
983	575	462	54.5	101.6	63.8	7.7	61.4
1153	580	470	54.7	104.5	71.2	6.8	59.5
1173	580	470	54.0	103.8	69.8	7.0	59.8
1196	580	470	53.5	101.2	63.2	7.8	61.6
1221	579	460	54.0	98.5	71.3	7.0	58.0
1245	575	462	54.0	100.8	70.7	7.0	58.8
1279	585	470	54.1	99.2	60.8	8.0	62.0
1293	585	470	55.0	99.2	67.9	7.1	59.4
1365	602	405	54.0	58.5	67.5	7.5	46.4
1387	565	385	53.5	64.1	82.5	6.6	43.7
1392	570	395	54.0	67.4	81.0	6.6	45.4
1539	570	405	54.0	63.1	44.5	11.9	58.6
1549	590	430	54.5	97.4	44.8	11.2	68.5
1573	598	440	54.0	93.3	46.6	10.7	66.7
1597	604	440	54.0	88.6	49.1	10.1	64.3
1622	595	380	54.5	65.6	53.2	9.9	55.2
1645	591	411	54.5	75.8	49.9	10.2	60.3
1668	591	410	54.0	75.1	53.6	9.6	58.4
1690	569	400	53.0	77.2	54.9	9.9	58.4
1718	575	408	54.0	78.5	63.2	8.3	58.4
1740	580	410	54.0	76.4	68.4	7.6	52.7

[1] Normalized to 600 psi - 68°F

TABLE II - 1 (Continued)

TUBULAR RO SYSTEM OPERATIONAL DATA 310 MODULES

Phase III & IV - 15 Modules (continued)

Elapsed Time (Hrs)	Pressure In psi	Pressure Out psi	Temp °F	Rate sec/gal	Product Rate sec/gal	Flux[1] gsfd	Product Water Recovery %
1763	592	410	54.0	72.6	69.9	7.4	50.9
1790	603	412	54.0	72.0	63.8	8.0	53.0
1814	604	418	54.8	72.4	71.3	7.0	50.3
1868	600	419	54.8	72.9	50.1	10.0	59.3
1884	600	420	54.0	74.6	49.7	10.1	60.6
1908	604	423	54.0	73.9	50.5	9.9	58.4
1932	610	438	54.2	78.7	55.5	8.8	58.6
1949	590	450	54.0	95.7	54.5	9.1	62.7

[1] Normalized to 600 psi - 68°F

TABLE II - 2

TUBULAR RO OPERATIONAL DATA 610 MODULES

Elapsed Time Hrs.	Pressure In psi	Pressure Out psi	Temp °F	Brine Rate sec/gal	Product rate sec/gal	Flux[1] gsfd	Product Water Recovery %
6	600	520	55.0	63.2	63.5	21.4	49.9
16	600	525	54.3	66.4	84.4	16.8	44.9
41	600	530	56.0	69.9	84.1	15.7	45.4
69	610	540	55.0	71.5	83.3	15.9	46.2
93	620	550	55.0	66.5	82.8	15.7	44.5
113	620	550	55.0	65.5	84.6	15.4	43.7
132	610	540	55.0	67.2	83.6	15.8	44.6
154	610	540	55.0	70.2	84.8	15.6	45.3
178	610	540	55.0	71.0	88.4	15.0	44.5
203	620	550	55.0	69.5	87.3	14.9	44.3
226	620	560	54.5	79.2	87.0	14.9	47.7
239	595	535	54.5	73.2	89.9	15.1	44.9
249	595	540	54.0	69.2	92.3	14.8	42.8
271	600	545	52.5	81.2	94.6	14.6	46.2
307	600	535	53.5	70.3	94.9	14.5	42.6
377	612	560	55.0	80.5	96.7	13.4	45.4
400	642	578	55.0	68.4	94.4	13.2	42.0
424	681	575	55.0	72.6	95.6	13.1	43.2
444	630	575	54.0	75.8	101.4	12.7	42.8
466	612	562	54.0	63.7	107.6	12.2	37.2
515	625	565	54.0	69.5	103.5	12.6	40.2
543	625	565	55.0	68.1	102.4	12.5	39.9
562	625	565	54.0	72.0	104.9	12.4	40.7
586	625	565	54.0	72.4	105.2	12.4	40.8
610	620	565	54.0	67.2	106.8	12.2	38.6
624	620	565	54.0	73.8	90.3	13.5	43.3
671	625	565	54.5	74.1	91.3	13.2	43.2
700	625	570	55.0	76.3	102.0	12.5	42.8
720	615	560	54.5	72.2	105.2	12.4	40.7
746	624	570	55.0	74.0	103.2	12.3	41.8
768	630	570	55.0	71.3	102.0	12.4	41.1
794	625	570	54.0	70.5	105.0	12.3	40.2
797	605	535	54.0	59.0	106.8	12.7	35.6
807	605	535	54.0	66.0	106.8	12.7	38.2
810	615	570	54.0	66.0	76.2	17.1	46.4

[1] Normalized to 600 psi - 68°F

TABLE II - 3

FIELD ANALYSIS DATA TUBULAR RO SYSTEM

Hour Clock Reading Hrs.	Iron Total mg/l	Iron II mg/l	SO_4 mg/l	Total Hardness Ca CO_3 mg/l	Calcium Hardness Ca CO_3 mg/l	pH Units	Meter TDS mg/l
			BRINE ANALYSIS				
1253	360	270	1900	2300	1340	2.9	3750
1282	300	260	3000	3400	1360	2.9	4350
1301	405	305	2700	2380	1380	2.8	4400
1328	308	186	1900	2180	980	-	3400
1348	540	310	2800	2720	1460	2.9	4100
1380	310	270	2300	2400	1300	3.1	3950
1445	215	210	1800	1800	800	3.0	3400
1496	172	160	1800	-	-	2.8	2425
1632	145	130	-	1400	700	3.0	3000
1682	199	182	2500	2100	1000	2.9	2250
1778	112	95	1180	1000	530	3.1	1950
1826	82	88	1360	1010	540	3.1	1950
1938	103	109	1440	1170	500	-	2000
1956	109	100	1560	1000	500	-	1900
2003	110	102	1400	1170	510	-	2100
2051	102	99	1240	1150	520	-	1950
2717	230	200	1500	1500	750	3.1	2900
3140[1]	200	160	1450	1500	500	-	2000
			PRODUCT WATER ANALYSIS				
1253	1.92	1.72	23	20	15	4.4	38
1282	1.92	1.82	23	29	19	4.5	39
1301	1.96	1.78	23	31	22	4.3	38
1328	1.78	1.76	28	35	15	-	35
1348	1.94	1.76	27	60	20	4.3	41
1380	1.76	1.67	21	18	13	4.7	36
1445	1.71	1.62	14	16	10	4.6	32
1496	1.59	1.48	14	16	7	4.4	33
1632	0.85	0.70	5	5	3	4.3	20
1682	0.35	0.32	1	2	1	4.3	20
1778	0.89	0.83	9	8	6	4.6	22
1826	0.90	0.88	12	16	9	4.5	21
1938	1.05	0.98	9	13	7	-	21
1956	1.00	0.99	11	11	8	-	21
2003	1.04	0.88	13	13	5	-	23
2051	1.08	1.05	12	13	7	-	22
2717	1.39	1.35	17	14	7	4.3	38
3140[1]	0.65	0.62	2	6	2	-	10

Note: See Hollow Fiber Data Appendix I for Raw AMD Analysis

[1] This Data for 610 Modules, all other Data for 310 Modules

TABLE II - 4

LABORATORY ANALYSIS DATA TUBULAR RO UNIT

Hour Clock Reading (hrs)	Ca mg/l	Mg mg/l	Mn mg/l	Total Fe mg/l	Si mg/l	Al mg/l	TDS mg/l
			Brine				
1229	383	295	44.8	238	18	25.5	4224
1301	430	337	52.0	288	19	25.0	4753
1396	384	310	48.0	256	19	27.0	4307
1591	212	167	27.6	123	15	17.0	2299
1898	205	149	47.4	115	15	17.0	--
2051	198	147	25.0	114	15	16.1	2334
2528	194	138	25.3	129	14	15.0	3226
3162[1]	232	170	28.2	152	20	15.9	2616
3489[1]	168	132	21.7	102	22	14.0	2074
			Product Water				
1229	3.2	1.9	0.36	1.7	7.0	0.2	56
1301	4.5	2.1	0.39	1.8	7.7	0.3	49
1396	1.9	1.9	0.36	1.6	6.0	0.3	63
1591	2.7	1.7	0.29	1.3	7.0	0.2	43
1898	2.2	1.1	0.24	0.9	--	--	--
2051	2.2	1.2	0.25	1.1	7.1	0.2	68
2528	2.7	1.5	0.28	1.6	8.0	0.1	41
3162[1]	0.7	0.5	0.08	0.5	1.0	0.2	33
3489[1]	0.5	0.4	0.08	0.3	1.0	0.1	17

NOTE: See hollow fiber data Appendix I for raw AMD analysis
[1] This data for 610 modules, all other data for 310 modules

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6 Title

REVERSE OSMOSIS DEMINERALIZATION OF ACID MINE DRAINAGE

10 Author(s)

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27 Abstract

The objective of this study was to determine the operational methods and procedures necessary to successfully demineralize acid mine drainage utilizing reverse osmosis (RO). Phase I consisted of laboratory bench scale investigations to determine methods for controlling iron fouling and to select a process flow sheet. Phase II was field operation based on the flow sheet selected in Phase I.

The field test site was located in Mocanaqua, Pennsylvania. The source of acid mine drainage was the discharge from an abandoned underground anthracite coal mine. Treatment prior to RO consisted of filtration (10μ) followed by ultraviolet light disinfection. The brine from the RO unit was treated by neutralization, oxidation and settling. The field test phase spanned a four month period. Frequent samples were analyzed to characterize the operation of the system.

The results obtained indicated that it was feasible to demineralize acid mine drainage by reverse osmosis. Membrane fouling due to iron was satisfactorily controlled. The recovery of product water was limited to about 75% due to calcium sulfate fouling. Product water was of potable quality in all respects except for iron, manganese, and pH. Neutralization, oxidation and filtration would be required to meet potable standards.

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